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AN APPRAISAL OF NEUTRALIZATION PROCESSES TO TREAT COAL MINE DRAINAGE

Harold L. Lovell

Pennsylvania State University

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Four different quality drainages were treated and detailed results tabulated. Appropriate unit operations, from water collection to sludge disposal were considered for eight different reagents (calcium/magnesium carbonate, oxide and hydroxide; caustic soda and soda ash). Necessary process variations were possible with a versatile 500,000 gpd facility.

Limestone has the least cost per neutralization equivalent and may be used with drainage containing up to 500 mg/l iron II. The resulting sludge is dense, rapid settling and dewaterable. Lime can treat any drainage efficiently but may result in excess consumption. Lime-produced sludges are voluminous and difficult to handle. Unitized sludge recycle process, using lime, forms a dense, dewaterable sludge. Iron II oxidation was accomplished by air in alkaline systems or with autotrophic bacteria in acid system.

Thickeners are preferred to settling lagoons to separate sludge in larger plants. Dewatered sludge requires least disposal volume. Drying basin dewatering is pragmatic, but filtration or centrifugation are effective, but more costly. Settled sludge may be discarded directly to abandoned deep or surface mines; while dewatered sludge may be placed in landfill areas. (EPA)

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AN APPRAISAL OF NEUTRALIZATION PROCESSES TO TREAT COAL MINE DRAINAGE

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ABSTRACT

The treatment of four different coal mine drainage waters was studied utilizing neutralization processes. Each of the unit operations, from water collection to sludge disposal, was considered by drainage reaction with eight different reagents. The process variations required by each reagent was possible with the specially-designed 500,000 gpd facility.

Limestone has the lowest delivered cost per neutralization equivalent and may be utilized with waters containing up to 500 mg/l iron II, which includes most coal mine drainages. A rapid settling, dewaterable sludge is formed. Lime can treat any drainage with efficiency, but requires control to avoid excess consumption by reaction with aluminum and magnesium. Lime forms a voluminous sludge which is difficult to handle. Iron II oxidation may be accomplished with air in alkaline systems or via autotrophic bacteria in acidic systems. Unitized sludge recycle process forms dense, dewaterable sludge from highly mineralized waters.

Thickeners are preferred to settling lagoons for sludge separations in larger plants. Sludge dewatered by drying basin, filtration or centrifugation requires the least land-fill disposal area. Alternatively, near-by abandoned deep or surface mines may be advantageously used for settled sludge disposal.

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SECTION I

CONCLUSIONS

Any of the coal mine drainages (CMD) considered in these studies may be treated by neutralization processes to meet State specifications in Pennsylvania: e.g., maintain a pH between 6.0 and 9.0, have less than 7 mg/l iron, and have no titratable acidity. Most of the aluminum content of the waters is also removed and under some conditions, there is a reduction in sulfate, magnesium, and manganese content. The product waters usually have no appreciable decrease in total dissolved solids and contain increased cation concentrations of the neutralizing reagent. The treated waters may be utilized where a "hard" water is suitable. They support aquatic life and are much less pernicious than the original mine drainages.

The treatment may be accomplished using any of the reagents studied - provided adequate process conditions are established. However, the efficiency of reagent utilization, the unit treatment cost, the required processing steps and the quality of the water product will vary widely among the several reagents.

Each treatment plant must be individually designed. Neutralization reagent selection, based on feed water quality, flow levels, and site conditions, is logically the initial decision. Reagent selection should also be justified by its process compatibility, reliable availability, transportation considerations, and cost at the proposed plant location. Continuous treatment capability must be insured.

The neutralization reagent determines the required unit operations and many of their parameters, thus the basic plant design.

The reagent limestone has the lowest delivered cost of those considered. This value approximates 1.92×10^{-6} cents per gallon per mg per liter acidity which is less than 40 percent of the next higher cost reagent, calcined lime. Excepting dolomite, limestone has the slowest reactivity with mine drainage and is most subject in its response to physical parameters as particle size and hardness. The physical parameters establish the mode of limestone utilization and ultimately its efficiency of use. The CMD reaction with limestone forms bicarbonates, carbonates and carbonic acid whose concentrations must be controlled to allow the pH of the water to rise. This latter response may limit selection of limestone in respect to the iron II concentration of the mine water. Waters containing up to 100 mg/l Fe II may be effectively treated using limestone without separate iron oxidation in the simplest system. Waters with Fe II contents between 100 and 500 mg/l also may be effectively treated with limestone but require prior iron oxidation. At still higher iron II levels, oxidation rates may limit pragmatic treatment. Preneutralization oxidation iron II extends the limits of limestone application, as does post-limestone neutralization additions of hydrated lime slurry. Most known mine waters may be effectively treated with limestone. Overtreatment using limestone is impossible. A dense, rapid-settling, dewaterable sludge results. Reagent usage efficiency can exceed 80 percent.

Hydrated and calcined lime are comparable reagents and are the least expensive reagents capable of treating any coal mine drainage in a feasible system. They react rapidly. When these reagents are properly reacted and controlled, nearly perfect reagent utilization efficiency is attainable. However, difficulties in control of reagent addition increase with the acidity of the water. The pH of the lime-treated water defines lime consumption and iron II oxidation rates. The pH-lime consumption relationship in these waters is complex since further reactions with aluminum and magnesium may occur. Overtreatment easily results using calcium hydroxide and may lead to unacceptable effluents and very high reagent costs. These reagents usually form a voluminous, gelatinous slurry of the impurities which settles to a low density sludge. These sludges gel on aging. They have poor handling and dewatering characteristics.

Hydrated lime may be reacted as a solid or made into an aqueous suspension, whose concentration is a process variable. The latter form is preferable.

Pebble-sized calcined lime should be slaked prior to treatment with the water. Although less expensive per unit neutralization equivalent, calcined lime is recommended only when consumption levels exceed 25 tons/week. Lower levels make slaking control difficult due to available equipment size.

Dolomite and its manufactured products perform similarly to high calcium reagents, except they are usually more costly and less available. They exhibit no apparent advantage in sludge volume or characteristics, except with excessively polluted waters. In that case, calcium sulfate deposition within the process system may be more controllable, but at the sacrifice of a less acceptable process effluent. Dolomite is less reactive than limestone, primarily due to its greater hardness. This low reactivity limits its application to waters which are very lightly mineralized. These reagents may lead to undesirable increases in magnesium content of the treatment effluents.

Sodium hydroxide is utilized as a solution and offers mechanical advantages. It is the most reactive, expensive and dangerous reagent to utilize. Its cost can vary widely with reagent physical form, with treatment plant location, and with transportation considerations. Sodium hydroxide is even more sensitive to the control difficulties cited for calcium hydroxide slurries and is also capable of near-perfect use efficiency. The resulting sludge characteristics are usually equivalent, but may be less desirable than those formed with calcium hydroxide. The sludge may contain less calcium sulfate and the plant effluent may be considerably different than with lime reagents. Sodium hydroxide is especially desirable as a short time, emergency control reagent to prevent unacceptable or accidental discharges.

Sodium carbonate, in the same neutralization cost range as solid sodium hydroxide, is utilized as a solution prepared prior to reaction. Although of high reactivity, the ensuing equilibria involving carbonates, bicarbonates and carbonic acid require decarbonation to allow the pH of the waters to rise and reactions to be completed. This reaction complication, which is similar to that exhibited by "naturally-alkaline coal mine drainage," when added to those difficulties cited for calcium hydroxide slurries, makes reaction control in the use of this reagent difficult. At least 100 percent excess reagent consumption can be anticipated. In properly reacted systems, the sludge characteristics may be somewhat superior to hydroxide reagents.

The mode of alkali addition and its reaction with the feed mine drainage waters, as established by the reaction control procedure (as a pH probe and its location in the system), are prime factors in determining reagent utilization efficiency.

A process was studied which recycles the impurity sludge. It provides an effective procedure to treat coal mine drainage by developing a crystalline, dense sludge. This process reduces sludge volume. The procedure is most applicable to highly-mineralized waters having a high iron II content. Sludge disposal problems are usually extensive for such waters. The process holds major advantage in situations where minimum plant space is available. The final sludge product, resulting with lime reagent, has densities in excess of 16 percent solids by weight. The sludge is desirably responsive to dewatering by vacuum filtration or sludge drying basin.

Separate air oxidation of iron II (at levels above 100 mg/l) is nearly essential with most of these reagent systems. It is preferably performed in an alkaline system. Procedures to estimate the rate of this oxidation are given and show that the oxidation rate dependency involves pH, temperature, and concentrations of iron II and aluminum. Inadequate retention times for this oxidation are more critical than oxygen transfer rates which are readily attained. The characteristics of this reaction, especially the changing rate with iron II concentration, requires its careful integration within a plant design.

Biochemical oxidation of iron II in an acid system, prior to neutralization, is feasible and extends the range of waters to which limestone as a reagent may be applied. In addition to the reaction parameters cited for air oxidation, the biochemical system requires consideration of cell concentrations and growth rates as controlled by cell metabolism and interferences. The surface biochemical reactor permits oxidation rates (in thousands of mg/l/hr) which are many fold faster than deep vat systems. The surface reactor system is easy to control and has little or no operational costs while the deep vat system has undesirable reaction rate limitations and is more subject to interferences. Further study of the surface biochemical reaction regarding temperature and sludge deposition is desirable.

The formation of large volumes of low density impurity sludges may be minimized to realistic levels during the initial water treatment. Limestone can be used to treat waters with iron II concentrations of less than 500 mg/l while more severely polluted waters may be effectively treated using lime in a sludge recycle system to yield more desirable sludge levels.

Sludge settling rates are about 2 ft/hr but vary widely with feed water quality, reagent choice and mode of utilization. These settling rates may be increased by addition of high molecular weight, polymeric, weakly anionic flocculants.

Either a settling lagoon or thickener may be satisfactorily utilized for solid-fluid separation of impurity sludges. As plant capacities exceed 300,000 gpd, thickeners become increasingly favored from both operational and economic bases.

Settling lagoons become filled with sludge. It is difficult and costly to remove the settled sludge. Procedures to implement sludge removal from lagoons or obviate it, increase costs favoring the use of thickeners.

Observations indicate that bottom-fed sludge solid-fluid separation techniques utilizing limestone-produced sludges can drastically reduce size requirements for settling lagoons and thickeners.

Secondary sludge dewatering is usually indicated. In some cases, economics favors discarding primary settled sludge in abandoned deep mines or surface strip pits.

Secondary sludge dewatering may be accomplished by sludge drying basin, filtration or centrifugation to form a product handleable as a solid containing 15-50 weight percent solids. The sludge drying basin has negligible operating costs; centrifugation is versatile in application while filtration has performance advantages.

Effectiveness of these unit operations increases with feed solids content. Vacuum precoat filtration is economically feasible only when solids production rates exceed an attainable 100 lbs dry solids/ ft^2/day. Filtration may be the only approach when the filtrate has a specified use.

Top loading filtration techniques may be less expensive since thicker cakes are achieved prior to blinding.

Freezing of wet, dewatered sludge modifies its physical structure to accelerate further dewatering and volume reduction upon thawing.

Movement of dewatered sludge to a "land-fill" disposal area and continued outdoor drying can result in a further volume reduction (two-thirds). This provides minimal disposal requirements. The dried sludge may be further mechanically compacted, graded and seeded.

SECTION II

RECOMMENDATIONS

The biochemical-iron II oxidation-limestone neutralization treatment system should be studied in continuous, year-round operation to improve the reliability and extend the applicability of the process. The areas to be emphasized:

A. Biochemical-iron II oxidation
   1) Temperature control under winter conditions
   2) Seek rates in deep vats equivalent to the surface-type reactor
   3) Cleaning surface-type reactor media and/or minimize sludge deposition.

B. Further quantify performance and seek increased capacity of rotary limestone reactor.

C. Bottom-feed solid-fluid separations to confirm potential reductions in thickener or settling lagoon areas indicated in these studies.

D. Top-feed filtration of impurity sludges.

E. Evaluate changes in sludge drying basin design to improve drainage rates and to establish optimum usage cycles.

Study treatment of extremely mineralized mine waters with sulfate contents over 3000 mg/1, seeking to minimize gypsum deposition by considering the use of mixed reagents (magnesium and sodium).

SECTION III

INTRODUCTION

This study concerns treatment of coal mine drainage (CMD) waters by neutralization processes. Such drainage creates an unacceptable condition in receiving streams to an ecologically-conscious society. Major portions of the river systems of coal-producing states, especially those with long histories of coal mining, as Pennsylvania and West Virginia, are extensively polluted. There are currently no feasible means to stop this water contamination, although a number of actions are possible to reduce the volume and extent of this pollution loading.

Pennsylvania and other states have, in recent years, enacted regulations which establish specifications for minimum water quality that may be discharged from operating coal mines. The goal is to stop the environmental degradation at the present levels and alleviate the pollution through continuing programs. In most cases the existing specifications can be met only by treatment of the water discharged from coal mines. In some instances, state action programs have been initiated to correct such discharges from abandoned coal mining operations.

Although concern and research related to such treatment has been evidenced for nearly fifty years, there has been inadequate practical engineering experience and knowledge which can be efficiently, effectively, and economically applied. The operators of existing facilities believe their operational costs are excessive. Attempts by industry and government to design and operate treatment facilities to meet the regulations have frequently resulted in frustration and have developed many questions.

The Pennsylvania Coal Research Board in December, 1965, requested The Pennsylvania State University to acquire the necessary data to permit effective, economical treatment. This information was to be developed from a small, but full-scale treatment facility utilizing the prevalent neutralization process concepts. This report presents the observations from that operation.

The research program, which was subsequently cosponsored by the Environmental Protection Agency, encompassed design and construction of a mine water treatment plant capable of processing 500,000 gallons of water per day. The plant was located at Hollywood, Pennsylvania (Clearfield County). Supporting laboratory studies were made at the University Park, Pennsylvania campus of The University. The experimental plant would be suitable for obtaining cost data and amenable to experimentation with innovations. The results of plant operation were to indicate performance and costs, including an evaluation of process elements.

The plant design was based upon treatment of the four available drainage waters, each with different characteristics. It provided capability to utilize any feasible alkali. The unit operations included ranged from water collection stations through the treatment steps to sludge disposal. The design versatility permitted interconnection between these unit operations to consider several processes including: (1) The "Yellowboy Process" (lime-air oxidation) with provision for solid-fluid separation by either thickener or settling lagoon, (2) Biochemical iron oxidation-limestone neutralization process with the oxidation achieved in either deep tanks or in a surface-type reactor, and (3) Recycle sludge process. Dewatering procedures applicable to the settled sludge sought to evaluate potential methods to control this troublesome and costly aspect.

Neutralization processes are designed to produce a water which will have minimum detrimental effects on the bioecological system. The treated waters must not detract from the appearance of receiving streams and be acceptable for navigational, recreational and industrial purposes. The development of potable-quality water for human consumption has not been a primary objective. These processes adjust the pH of the waters to levels above 6.0. This treatment creates conditions which form insoluble compounds from the iron, aluminum and sometimes calcium sulfate contained in the drainage. The insolubles must be removed from the treated water. The quality criteria for such discharged waters established by the Sanitary Water Board of the Commonwealth of Pennsylvania in 1964 have served as general guides. These require that discharges to the streams of the Commonwealth have: (1) a pH between 6.0 and 9.0, (2) iron content less than 7 mg/liter, and (3) no titratable acidity. The Hollywood plant was operated to meet these specifications.

It appears these criteria should result by reacting any inorganic alkali providing it does not introduce other inimical components. The volume of such waters which may require treatment reaches millions of gallons daily, thus pragmatic considerations also necessitate the utilization of processes having minimal requirements for capital equipment, labor, process control, and operating costs. They must also offer a high degree of reliability.

These processes can be conceived as cation exchange reactions in which the hydrogen, iron, and aluminum ion species in the drainage waters are replaced by calcium, magnesium, or sodium ions. The term "cation exchange reactions" describes the replacement of the undesirable cation components of the drainage waters by more acceptable soluble ions. The detrimental components are converted to insoluble substances. In no way does the term refer to processes labelled "ion exchange" which involve a second phase, insoluble substrate. The water products resulting from neutralization treatment are not significantly improved in total dissolved solids content.

Other processes, described as "demineralization procedures," do result in higher quality water products (low dissolved solids). They are more expensive and of limited application.

Although the number of neutralization processes that can be used to achieve these minimal quality standards is limited, the factors involved in choosing a process are more complex than commonly conceived. Once the characteristics of the product waters are defined, the process choice depends upon the loading of the drainage to be treated, its volume, logistical problems of the water sources and the treatment facility as relates to each potential reagent. The chemistry of the reactions defines the engineering criteria. The cost and availability of the various reagents establishes the pragmatic choice. The study seeks an insight into these various determining factors.

Since waters dissolve various substances as they flow through coal measures, the characteristics of the drainage differs widely with strata mineralogy. The principal mineralization results from the soluble oxidation products of the indigenous mineral pyrite. The components of concern include the following cations: hydrogen, ferrous, ferric, and aluminum. Sulfate is the predominating anion. The species in which these ions exist in coal mine drainage waters probably vary but they have not been fully defined. Furthermore, under the *in situ* conditions, these components are undergoing continuing internal reactions. These reactions may be enhanced biochemically or by their contact with other strata, adding ions such as calcium, magnesium, sodium, potassium, carbonate, bicarbonate, and, to a lesser degree, manganese and chloride as well as a host of other ions at trace level concentrations. The many interactions and equilibria which exist in mineralized waters or may be developed with the introduction of inorganic alkalies create systems that may be described as "buffered." The waters considered in this study, whose analyses are detailed in Appendix A, are reasonably typical of coal mine drainage Classes I and II[1a]. They represent the preponderance of coal mine drainages.

The generalized reactions which occur in the desired cation exchange processes are of two types:

1) Neutralization - The available hydrogen ions in the drainage waters react with negatively-charged radicals of the reagents to form weakly-ionized species, such as water. In this chemical system, the solubility products of the iron and aluminum ionic species are exceeded, thus reducing their solution concentrations.

2) Oxidation of iron II to iron III - The solubility product of ferrous hydroxide is much larger than that of ferric hydroxide. With the predominance of the ferrous species in most drainage waters, the provision of a greater hydroxyl concentration is necessary than required for ferric systems to achieve acceptable levels of iron concentration. The final pH of waters so treated may exceed the criterion for hydrogen ion concentrations. During iron II oxidation, additional acid species form which represent an unacceptable pollution capability. Oxidation of the iron to the higher valence during processing obviates these problems. The iron II oxidation with oxygen from air tends to proceed after neutralization. The oxidation rate increases with pH at levels above 4.0.

NEUTRALIZATION

In these neutralization reactions, the hydroxyl ion is really the only ion species feasible to form a weakly ionized, polar compound with the hydrogen ion. The carbonate ion will react with the hydrogen ion to form carbonic acid, which will dissociate to form water and dissolved carbon dioxide. Because of the limited solubility at ambient temperatures and pressures, the carbon dioxide can be removed from the water. The carbonates also form insoluble compounds with iron and aluminum species. Thus, the two anions, hydroxide and carbonate, are the fundamental reagent constituents in the treatment of coal mine drainage waters. They are available as compounds of calcium, magnesium, sodium, and ammonia at costs feasible for these processes. Other cations form hydroxides and carbonates, but their cost and availability eliminate them from consideration.

Ecological principles usually preclude the use of ammonium hydroxide since the soluble ammonium ion serves as a nutrient for plant and animal organisms whose growth in receiving streams is undesirable. The physiological action of magnesium ion as a cathartic, especially when associated with sulfate ion, makes its introduction into streams undesirable.

The reagents which may appropriately be considered for achieving the cation exchange reactions necessary for neutralization of mine waters are limited. These include: calcium carbonate (high calcium limestone), calcium oxide (calcined, quick, or pebble lime), calcium hydroxide (hydrated lime), calcium carbonate-magnesium carbonate (dolomite or dolomitic limestone), calcium oxide-magnesium oxide (burnt or calcined dolomite), calcium hydroxide-magnesium oxide (hydrated dolomite or partially hydrated dolomite), calcium hydroxide-magnesium hydroxide (pressure hydrated dolomite), sodium carbonate (soda ash), and sodium hydroxide (caustic soda). The oxides of calcium and magnesium meet the requirements for reagent anions (e.g. they function as hydroxides) since as anhydrides they react with water to form hydroxides.

Pertinent data regarding these reagents are listed in Table 1; while their costs, as experienced during the project studies, are cited in Table 2. The chemical and particle size analyses of the reagents used in these studies are given in Table 3.

Many other considerations become basic in choosing among these reagents. In planning a treatment system, the selection of a reagent would be the initial decision since this information forms the basis of the processes, plant design detail, and economics. The selection is established by the drainage water analyses and volume, plant location and the reagent performance potential. These studies were designed to provide, from operating experience, the criteria for the reagent performance potential. Each of the following factors should also be optimized for the reagent: source geographical distribution, availability, transportation, handleability, cost, reactivity, chemical and physical properties and their mechanistic action within the treatment system as this affects the quality of the final water product and the characteristics of the impurity sludges.

High calcium and dolomitic limestone are naturally-occurring mineral products. The remaining reagents are all manufactured products which partially accounts for their higher cost. Quality and reactivity of high calcium and dolomitic limestone and their products vary widely with origin and manufacturing process[81]. Calcined materials contain five percent or more inert constituents which effect costs and processes. The differences in chemistry between dolomite and high calcium materials must also be considered. The physical form of caustic soda[95] and soda ash (solid, flake, solution, etc.) selected must be carefully evaluated. There are major cost differences and they have an impact in determining reaction conditions, ambient storage and reaction temperatures, transportation, handling, storage, safety, etc. Since the demand for these reagents for mine drainage treatment is in no way market controlling, consumption for this purpose must be planned around prevailing commercial products; e.g. particle sizes and qualities. The various reagents have definite limits of interchangeability within a given plant design.

By-products of manufacturing processes, as "carbide lime" from acetylene manufacture, may offer cost or process advantages but

Table 1. MINE DRAINAGE TREATMENT REAGENT DATA

Reagent	Formula	Formula Weight	Theoretical Neutralization Equivalent lb/gallon/mg/l acidity x 10^{-6}
Limestone	$CaCO_3$	100.09	8.3453
Calcined lime	CaO	56.08	4.6758
Hydrated lime	$Ca(OH)_2$	74.10	6.1783
Dolomite	$CaCO_3 \cdot MgCO_3$	184.42	7.6749
Calcined dolomite	$CaO \cdot MgO$	96.40	4.6759
Hydrated dolomite	$Ca(OH)_2 \cdot MgO$	114.42	4.7701
Pressure hydrated dolomite	$Ca(OH)_2 \cdot Mg(OH)_2$	132.44	5.5213
Caustic soda	$NaOH$	40.01	6.6719
Soda ash	Na_2CO_3	106.00	8.8377

Table 2. REAGENT COST DATA

(1968-1971)

Reagent	$/ton fob source	Transpor-tation ¢/ton mile	Delivered Cost Hollywood $/ton	¢/1b	Reagent Cost at[a] Hollywood ¢/neutralization equivalent / ¢/gal/mg/L/ acidity x 10^{-6}	Delivered Cost Ratio to Limestone on Neutralization Equi-valent Basis
Limestone						
Stone sizes	1.50	3.	4.50	0.23	1.92	--
Waste	0.90	3.	3.90	0.20	1.67	0.87
Pulverized		3.	8.50	0.43	3.59	1.87
Calcined lime						
Bulk	16.00-18.00	3.	21.00	1.05	4.91	2.56
Bagged	19.50	3.	23.50	1.18	5.52	2.88
Hydrated lime						
Bulk	18.25-20.75	3.	23.75	1.19	7.35	3.83
Bagged	21.25-23.75	3.	28.25	1.41	8.71	4.54
Dolomite						
Stone sizes	1.80-2.50	4.	10.00	0.50	3.84	2.00
Pulverized	7.00	4.	15.00	0.75	5.76	3.00
Calcined dolomite						
Bulk	16.00	4.	24.00	1.20	4.82	2.51
Bagged	21.00	4.	29.00	1.45	5.83	3.04
Hydrated dolomite						
Bagged	19.75	4.	29.50	1.48	7.06	3.68
Pressure hydrated dolomite						
Bagged	23.00	4.	31.00	1.55	8.56	4.46
Caustic Soda						
50% Solution						
Tanker			66.00	3.30	22.02	11.47
Drum			268.00	13.40	89.40	46.56
Flake						
Drums, Bulk			78.00	3.90	26.02	13.55
Solid, small lots			110.00	5.50	36.69	19.11
Soda Ash						
Powder						
Bulk			40.00	2.00	17.68	9.21
Bags			76.00	3.80	33.58	17.49

[a] Assumes 100% reagent purity.

Table 3. ANALYSES OF REAGENTS USED DURING OPERATING PROGRAM
Chemical Analyses (Dry Basis)
Weight Percent

Reagent	Ca	Mg	SiO_2	Total Insolubles	Loss on Ignition	CaO Equivalent
Hydrated lime[a]	50.84	0.25	1.42	1.48	25.33	53.65
Calcined lime	68.61	0.60	--	1.35	1.08	96.00
Grit from slaking calcined lime	46.03	--	3.87	4.31	24.93	--
Limestone	38.79	0.19	1.08	1.86	42.88	47.40
Hydrated dolomite	29.58	17.33	2.23	2.40	26.43	
Calcined dolomite	38.77	23.57	2.62	2.68	2.28	
Dolomite	20.71	12.36	3.48	4.72		

[a] Partially recarbonated from atmosphere while in storage bin.

their use should be carefully evaluated, including consideration of quality, uniformity, long range availability, etc.

Pulverized limestone and lime products are commonly transported in bulk utilizing pneumatic handling. Loads range from 15 to 28 tons. Air pollution by dust release during unloading can be controlled and is to be avoided. Reagent and shipment costs by bags of up to 100 pounds capacity are higher and also increases subsequent reagent handling costs. Pulverized stone products, hydrated lime and calcined lime are conveniently stored in bins. Bin storage of hydrated lime can result in extensive water and carbon dioxide absorption within several weeks; thus long-term lime storage which can result from excessively large bins is to be avoided. In addition to bin storage, pneumatically-handled hydrated lime may be slurried (10-50 weight percent) directly from the delivery tanker and stored as a suspension. Larger particle sizes of limestone (> 3/8 inches) have the best handleability, do not require protected storage facilities, and are not affected by the weather. Some stone products having large amounts of fines may handle differently when wet.

The total components comprising reagent costs in mine drainage treatment extend beyond the fob unit cost (as per quantity, quality and bid), to incorporate transportation, handling, storage, stoichiometry, use efficiency, etc. The reagent cost may be expressed as cents/1000 gallons water treated and compared with the theoretical value. Overall one seeks the best system efficiency which is here defined as the ratio of total treatment costs to reagent costs. For the coal operator, a more meaningful expression may be cents/ton of saleable coal produced; or for the ecologist, some broader expression as $/mile of stream controlled.

Critical factors in reagent selection and use also include the location of their origin, quality, mode and points of manufacture-distribution, mode of transportation, and availability. Experience indicates a disinterest and a neglect of these factors, perhaps even intentional avoidance by researchers, government agencies, and engineers. A proper understanding of the role of these aspects is important to the correction of the coal mine drainage problem.

REAGENT REACTIVITY

The chemical characteristics of the reagents used to react with mine drainage are well known. Each of the reagents is subject to extensive dissociation in aqueous solution, as are the reactive ionic species in coal mine drainage. Consequently, the neutralization, "cation exchange" reactions occur essentially instantaneously when the reactive species contact. The reaction rate is primarily physical involving ionic species "availability" and "contact." Reagent and product solubility and mechanical aspects (mixing) thus really become the controlling factors. However, many different equilibria are involved in these systems and they play a major role in defining the direction and extent of the reactive system. The "relative" rate variations between the reagents are quite extensive as illustrated in Figure 1. Sodium hydroxide will be more responsive than hydrated lime, while soda ash can be expected to respond somewhat more slowly as controlled by the bicarbonate equilibria.

A review relating these reagents in the treatment of industrial waste acids (Lewis[2]) emphasizes the varied nature of the wastes (which incorporates coal mine drainage) and that they are not amenable to standardized treatment procedures. Lewis cites two considerations: the pH range over which treatment is to take place and the minimum time available for the reaction. For mine drainage, the first is controlled by the water characteristics and governmental regulations. The final pH is complicated by equilibria developed in the reactions. The time available for reaction is a pragmatic decision. Some reagents require very large reactors and power inputs as reaction times exceed ten minutes. Figure 2 illustrates the pH ranges of the treated waters which may be expected. The upper pH limit in carbonate systems is controlled by mechanical factors as particle size and conditions affecting the hydroxide-bicarbonate-carbonate equilibria.

FERROUS IRON OXIDATION

This reaction in mine drainage waters occurs in natural situations forming iron III as an unstable ferric sulfate. The subsequent hydrolysis forms various amorphorous and crystalline species of insoluble iron oxyhydroxides. Lovell and Lachman[3] have detailed these changes for a particular stream. The hydrolysis products, in wide ranges of composition, appear in the bottom of streams carrying coal mine drainage and are commonly referred to as "yellowboy". These deposits create the undesirable appearances of streams which identifies this pollution loading. This natural oxidation is attributed to slow air oxidation, to the action of indigenous autotropic bacteria, and possibly to some catalytic effects.

In the treatment of coal mine drainage with alkalies, it is usually considered desirable, if not essential, to convert the iron to the ferric state during the process. There are two reasons for this objective: (1) The solubility product of ferric hydroxide (1.1×10^{-36} at 18°C) is lower than that of ferrous hydroxide (1.64×10^{-14} at 18°C)[11a]. This means that the desired iron concentrations can be met at lower pH levels requiring the use of less alkali; it also provides for the potential use of limestone as a neutralization reagent, and minimizes the potential for side reactions occurring

Figure 1. Relative reaction rates for various alkali reagents

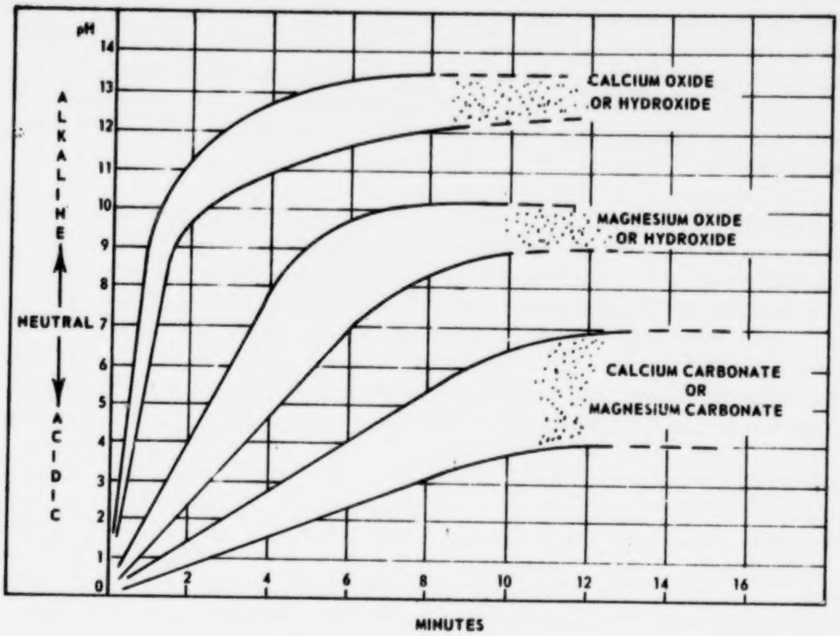
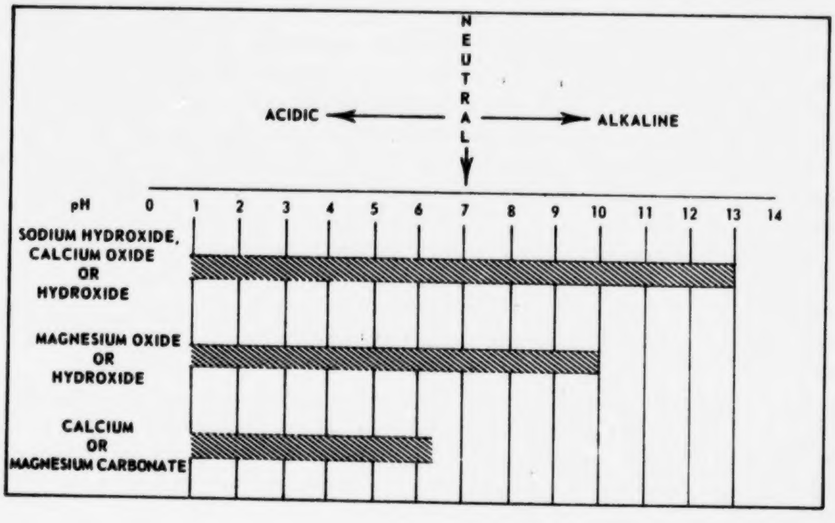


Figure 2. pH range resulting from various alkali reagents



with aluminum and magnesium species in the water which are favored at higher pH levels. (2) Iron II forms additional hydrogen ion upon oxidation. Thus, whether iron II exists in treated water or a sludge product, there is a potential for additional acid formation. Such continuing reactions also lead to process control problems and unstable systems.

The oxidation of iron II may be accomplished by atmospheric oxygen, by other reagents, as ozone, chlorine, peroxides and potassium permanganate, or by biochemical processes. A controlled process of iron II oxidation by oxygen from air is nearly universally utilized and has been the subject of considerable study. The process design for this oxidation is controlled by the two stages of the process: (1) The rate of iron II oxidation with oxygen, and (2) The rate of transfer of oxygen from air to the reacting system.

Rate of Iron II Oxidation with Air

The parameters, which have been established to govern this reaction are: temperature, pH, iron II concentration and aluminum III concentration. The rate increases proportionately with the first three but shows a reciprocal relationship to aluminum III concentration. There probably are other factors involved. Catalytic agents, which may increase oxidation rates, have not been shown to play a role in the reaction unless specifically added. Singer and Stumm[4] reported sulfate complexes may be inhibiting. Steinberg, et al.[100] describe increased rates when the reaction is carried out with high levels of Co-60 gamma radiation. Obviously, sufficient oxygen, presumably in the form of dissolved oxygen, must be available. This parameter is excluded from the rate considerations since it is assumed that the aqueous system will be maintained in a saturated condition by continuous air transfer. The oxygen concentration may be considered a constant at the prevailing temperature.

The overall reactions may be described by Equation 1 or as two steps in Equations 2 and 3.

$$4Fe^{+2} + O_2 + 10H_2O \rightarrow 4Fe(OH)_3 + 8H^+ \quad (1)$$

$$4FeSO_4 + O_2 + 2H_2SO_4 \rightarrow 2Fe_2(SO_4)_3 + 2H_2O \quad (2)$$

$$Fe_2(SO_4)_3 + 6H_2O \rightarrow 2Fe(OH)_3 + 3H_2SO_4 \quad (3)$$

Equations 2 and 3 are probably over-simplications of the actual mechanistic steps which occur. It has been reported[5] that a faster oxidation rate occurs if the initial iron II form is in the solid state $Fe(OH)_2$ as opposed to a soluble ionic species. This may be related to iron-sulfate complexes[4].

Potentially one mole of sulfuric acid is formed per mole of iron II oxidized and hydrolyzed. This acid product lowers the pH of the system and reduces the oxidation rate. Thus the equilibrium pH which develops in the oxidation stage becomes the parameter which controls the oxidation rate. The basic stoichiometry as shown by Equation 1 appears to prevail and requires 0.143 pounds of available oxygen per pound of iron II.

If the alkali is added prior to iron II oxidation, as in the "yellow-boy" process, the pH level must be adequate to neutralize the acid formed during oxidation and maintain an acceptable oxidation rate. Should a process involve post-oxidation neutralization, the effect of decreasing pH on the oxidation rate must be considered.

The experience with this oxidation reaction, as associated with various alkaline reagents, has been reviewed[1]. Stumm, et al.[4, 6, 7] have studied the reaction and reports the oxidation rate to be first order with respect to iron II and independent of iron III. The strong effect of pH is related to the role of the hydroxyl ion. Mihok[8] proposed a broad oxidation rate schedule, as a design base for limestone treatment systems:

Fe II Concentration mg/1	Oxidation Rate mg/1/hr
50	180
100	240
250	300
500	360
1000	450
1500	600
2000	750

pH range 6.5-7.5
Temperature 0-10°C

A detailed laboratory study of iron II oxidation rates, preliminary to these studies, was made by Stauffer and Lovell[5], under conditions anticipated in the operating treatment plant. The aqueous system was maintained saturated with oxygen from bubbling air, temperature was controlled and pH was maintained at a constant level by the automatic addition of sodium hydroxide solutions.

Stauffer's data, based on synthetic solutions which cover but a limited range of variables (pH from 4.6 to 5.7, iron II and aluminum concentrations to 800 mg/1 and 200 to 400 mg/1 respectively, and temperatures between 15 to 25°C), do not respond well to rigorous mathematical treatment. Even multiple linear regression treatment involving natural logarithms of temperature and concentration were non-significant. A reaction order can be established graphically. With the reaction rate being a differential of iron concentration with respect to time, a log-log plot yields a slope of 5/8 which can be expressed in integrated form:

$$(Fe_t^{+2})^{3/8} = (Fe_O^{+2})^{3/8} - k'T \tag{4}$$

$$\text{where } k' = 3/8k \ (-r = k[Fe^{+2}]^{5/8})$$

The levels of the constant k' are dependent upon the cited parameters and were related for Stauffer's data through the following multiple linear regression equation:

$$k' = b_O + b_1(pH) + b_2 \ln 1/T + b_3(Fe_O^{+2})^{3/8} + b_4 \ (1/Al) \tag{5}$$

where the coefficients are given in Table 4.

Table 4. REGRESSION COEFFICIENTS OF THE LINEAR RESPONSE SURFACE FOR THE RELATED RESPONSE OF IRON II OXIDATION RATES
(Based on Stauffer's Data)

Coefficient	k'
b_O	-5.44892
b_1	0.170686
b_2	-0.76211
b_3	0.013887
b_4	9.42902

Utilizing this constant k' with the initial and desired final iron II concentration, reaction retention time may be estimated as shown in Equation 6.

$$t_r = \frac{(Fe_O^{+2})^{3/8} - (Fe_t^{+2})^{3/8}}{k'} \tag{6}$$

where: t_r = Oxidation time in minutes

Fe_O^{+2} = Initial iron II concentration in mg/l

Fe_t^{+2} = Desired final iron II concentration in mg/l

k' = constant defined by Equation 5

T = temperature = 273 + degrees C°

pH = units

Al = initial concentration in mg/l

This procedure was tested statistically. The hypothesis was that the proposed model does not predict the rate constant. The fact that the analysis gives a significant result indicates that the probability of rejecting the hypothesis when it is true is less than 5 percent, but greater than 1 percent.

This reaction time estimation is minimum and is considered to be conservative due to the range limitations of Stauffer's data. This provides a safety factor when applied to design calculations. Table 3 in Appendix E lists various calculated rates for different waters. It is not recommended that this estimate procedure be employed with carbonate or bicarbonate systems due to the additional factors involved with the changing equilibria associated with such systems.

The design engineer should be especially cognizant of the rapidly decreasing oxidation rate with decreasing iron II concentrations. This situation is especially complex in a dynamic system where the final effluent iron II concentrations (which would be in equilibrium with the oxidation tank iron II concentrations) must be at minimal levels. The chemical air oxidation system is similar in this respect to the biochemical oxidation system discussed in Section XIII.

PROCEDURE FOR PLANT OPERATION

As an experimental facility utilizing waters from abandoned mines, there was no requirement to maintain a given effluent water specification. However, the operating control principle was to meet the water quality specifications of the Pennsylvania Sanitary Water Board. The Facility was operated under Sanitary Water Board Water Permit No. 267 MO 11 (April 28, 1967).

The alkali addition was controlled to maintain the required pH in the plant effluent based upon continuous measurement with a glass electrode system. The electrical signal from the pH transducer was utilized to operate an automatic flow control valve for the alkali slurry addition. For "yellowboy" process, the pH probe was initially located in the oxidation tanks which monitored the water near the end of the processing operations. Although the lag time was relatively long, the pH drift in the waters was not excessive. In latter tests the probe was located near the effluent overflow of the flash mixer to observe lag time response. Under these conditions there was a definite trend to introduce local excess alkali which led to unwanted reactions with aluminum and magnesium ions and consumes additional alkali. See Figure 7 and in Appendix B, Figures 9, 13 for these locations of the pH probe.

In the Densator operation, the pH probe was located external to the reactor but received a continuous sample stream from the secondary reaction zone. See Appendix B, Figure 14. There was no provision for automatic pH control with the limestone reactor although a pH probe was located in the effluent sump for continuous monitoring of the effluent.

The detailed plant operational program was based on a minimum of 48 hours operation under a given set of process conditions. It was impossible to follow the initial schedule due to uncontrollable conditions, including weather, stream flows, mechanical failures, and personnel limitations. Similarly, experience with the processes necessitated timing variations in some cases, since it was not feasible to operate within the prescribed limiting period, as in the Densator tests and biochemical iron oxidations.

The observations and data reported were developed over approximately one year's operating period. The operating personnel were usually limited to one man per shift. In addition to control of the treatment, they were responsible for sample collection, control analyses, repairs, and facility maintenance including janitorial services, snow removal and grass cutting.

SECTION IV

DESCRIPTION OF PILOT PLANT

The concept of the Hollywood Facility was developed by University personnel to evaluate and compare different chemical neutralization processes for treating coal mine drainage. Every effort was made to incorporate the concepts set forth in the Mine Drainage Manual of the Pennsylvania Sanitary Water Board[10a]. Certain unit operations would be less than total design capacity to reduce costs. Coal mine drainage waters would be pumped to the Facility, while gravity flow through the plant was sought with discharge of the effluent to natural drainage. The variability of the water sources and processes precluded optimum design for each situation. The basic unit operations would involve alkali neutralization, iron II oxidation, and solid-fluid separation of the impurity sludge. Provision included raw water storage conforming to Pennsylvania Sanitary Water Board criteria[10a] and sludge dewatering and disposal.

Capability would be included to utilize any alkaline reagent, including limestone which would require a specially-designed reactor. The thickener would involve a specially designed unit, the Infilco Densator, to permit study of sludge recycle concepts. The settling lagoon incorporated an original design for continuous sludge handling. Provision for the use of flocculant aids would be included. Automatic plant control would be limited to pH response of alkali introduction while instrumentation would permit data collection of several process parameters, including water quality, temperature, and flow. Power consumption would be measured on the larger motors. The summarized design plan is given in Table 5 and Figure 3. The initial design permitted evaluation of only one process at a time.

Based upon these criteria, Gannett, Fleming, Corddry and Carpenter, Inc., an engineering firm, prepared a basis for detailed design as related to the Hollywood site. The resulting data which formed the basis of design are presented in Tables 6, 7, and 8. Appendix B presents the detailed units and facilities planned. The estimated construction cost, based on these data, was $805,000.00. The actual costs are shown in Table 9. The power consumption observed for the various unit operations is given in Appendix B, Table 3.

A photograph of the resulting Facility is shown in Figure 4. In Figure 5 is presented the hydraulic profile for the various unit operations. Simplified drawings of the Facility are presented in Figures 6 and 7. The unit operation locations are indicated in Figure 6; the exterior piping in Figure 7. The thickener-Densator was constructed above

Table 5. BASIS OF DESIGN OF THE HOLLYWOOD FACILITY

Mine Drainage Water Volume		
gallons per day		500,000
gallons per hour		20,833
gallons per minute		347
Total Acidity - tons/day $CaCO_3$		2.08
Total Iron - tons/day Fe		0.21
Alkali Requirement		
tons/day, hydrated lime ($Ca(OH)_2$)[c]		1.62
Oxygen Requirement		
for iron oxidation pounds/hour[d]		2.52
Estimated Air Requirement		
ft^3/min, 45 minute retention		22
Thickener or Settling Pond Sludge Resulting from Process		
Solids content, percent by weight		1.0
gallons/1000 gallons water treated	a.	50
	b.	10
gallons/day	a.	25,000
	b.	5,000
gallons/hour	a.	1,040
	b.	208
Dry Solids in Sludge		
pounds/1000 gallons water treated		5
tons/day		1.25
pounds/hour		104
Densator-Thickener Dimensions (Diam. X SWD) feet		25 x 20
Drum Filter		
Solids loading pounds/square foot/hour	a.	5.0
	b.	15.0
1-6 foot diameter by 7 foot long rotary drum-belt vaccum filter Filter area, square feet		125

a. Based on sludge from normal settling.
b. Based on dense sludge.
c. Assumed to have 95% available CaO.
d. Assuming all iron in ferrous state.

Figure 3. Simplified schematic drawing of treatment plant

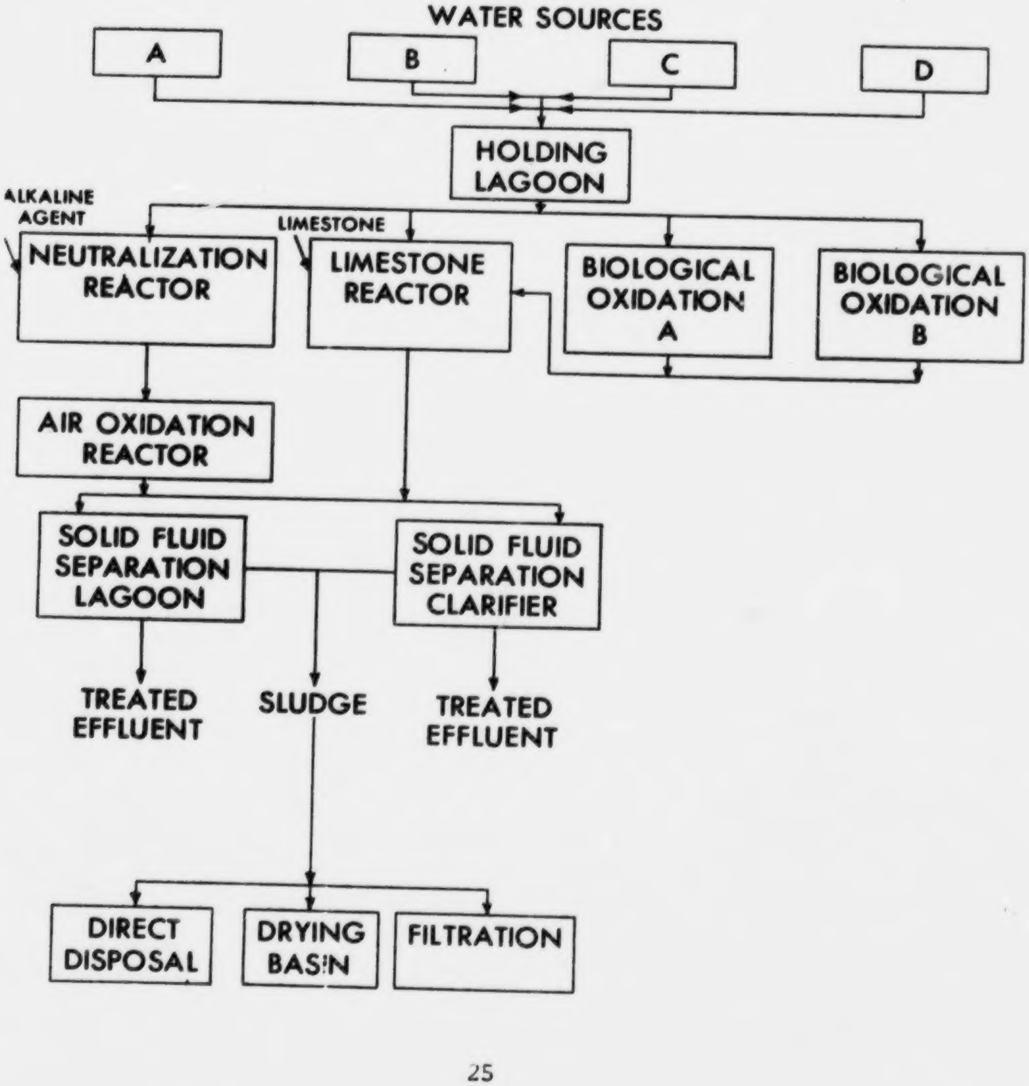

WATER SOURCES
A
B
C
D
HOLDING LAGOON
ALKALINE AGENT
NEUTRALIZATION REACTOR
LIMESTONE
LIMESTONE REACTOR
BIOLOGICAL OXIDATION A
BIOLOGICAL OXIDATION B
AIR OXIDATION REACTOR
SOLID FLUID SEPARATION LAGOON
SOLID FLUID SEPARATION CLARIFIER
TREATED EFFLUENT
SLUDGE
TREATED EFFLUENT
DIRECT DISPOSAL
DRYING BASIN
FILTRATION

Table 6. SOURCES OF MINE DRAINAGE ESTIMATED VOLUMES, CONSTITUENTS AND CHARACTERISTICS

Source	Proctor No. 2	Tyler Run	Proctor No. 1	Bennett Branch
Volume - mgd				
Maximum	0.063	0.600	0.288	not available
Average	0.025	0.119	0.199	30
Minimum	0.009	0.003	0.127	3
pH				
Maximum	3.81	4.30	3.49	4.31
Average	3.05	3.35	3.05	3.43
Minimum	2.70	2.91	2.60	2.80
Acidity (as $CaCO_3$) - mg/l				
Maximum	8,112	420	1,271	523
Average	6,700	330	780	278
Minimum	5,343	150	316	39
Total Iron - mg/l				
Maximum	3,604	90	199	105
Average	2,090	27	134	53
Minimum	1,516	2	99	11
Ferrous Iron - mg/l				
Maximum	2,300	90	199	105
Average	1,780	26	133	53
Minimum	1,360	2	99	10
Aluminum - mg/l				
Maximum	770	131	196	45
Average	725	100	141	30
Minimum	680	51	93	22
Calcium - mg/l				
Maximum	252	108	144	56
Average	150	77	102	40
Minimum	60	40	40	20
Magnesium - mg/l				
Maximum	129	114	132	46
Average	80	83	80	42
Minimum	3	58	10	39

Table 6 (continued). SOURCES OF MINE DRAINAGE ESTIMATED VOLUMES, CONSTITUENTS AND CHARACTERISTICS

Source	Proctor No. 2	Tyler Run	Proctor No. 1	Bennett Branch
Sulfate - mg/l				
Maximum	6,174	712	1,127	514
Average	5,150	640	1,080	385
Minimum	4,062	536	990	248
Temperature - °C				
Maximum	26	25	21	28
Average	19	16	15	18
Minimum	8	3	4	3
Specific Gravity	1.11	1.09	1.10	1.10

Table 7. TREATMENT PLANT DESIGN CONDITIONS

	Chemical Treatment	Biochemical Treatment
Daily Volume - million gal	0.500	0.100
pH	2.9	2.9
Acidity (as $CaCO_3$) - mg/l	1,000	1,000
Total Iron - mg/l	100	100
Ferrous Iron - mg/l	100	100
Sulfate - mg/l	1,100	1,100

Table 8. MAJOR TREATMENT PLANS

Plan I

a) Lime neutralization in flash mixer
b) Mechanical oxidation
c) Flocculation and settling in earthen lagoon
d) Overflow to Tyler Run
e) Waste underflow to vacuum filter, earthen sludge drying lagoon, and tank truck for hauling to local strip pits

Plan II

a) Lime neutralization in flash mixer
b) Mechanical oxidation
c) Flocculation and settling in Densator unit operated as clarifier or clarifier-thickener
d) Overflow to Tyler Run
e) Waste underflow to vacuum filter, earthen sludge drying lagoon, and tank truck for hauling to local strip pits

Plan III

a) Lime neutralization in Densator unit
b) Flocculation, settling and thickening in Densator unit operated on densating principle
c) Overflow to Tyler Run
d) Waste underflow to vacuum filter, earthen sludge drying lagoon, and tank truck for hauling to local strip pits

Plan IV

a) Biological oxidation by air diffusion, and turbine and surface aerators
b) Flocculation, settling and thickening of biological floc in Densator unit operated as clarifier or clarifier-thickener
c) Waste underflow to vacuum filter, earthen sludge storage lagoon, and tank truck for hauling to local strip pits
d) Limestone neutralization of Densator overflow
e) Flocculation and settling in earthen lagoon
f) Waste underflow to vacuum filter, earthen sludge storage lagoon, and tank truck for hauling to local strip pits

Plan V

a) Biological oxidation by rock filter
b) Flocculation, settling and thickening of biological floc in Densator unit operated as clarifier or clarifier-thickener

Table 8 (continued). MAJOR TREATMENT PLANS

c) Waste underflow to vacuum filter, earthen sludge storage lagoon, and tank truck for hauling to local strip pits
d) Limestone neutralization of Densator overflow
e) Flocculation and settling in earthen lagoon
f) Waste underflow to vacuum filter, earthen sludge storage lagoon, and tank truck for hauling to local strip pits

ground level to reduce installation costs and improve accessability to the unit. This required the influent to be pumped and created freezing problems by heat loss through the top and sides of the thickener.

The dimensions and volumes relating to the settling lagoon are given in Appendix B, Table 4. The lagoon discharge sump was equipped to record effluent temperature and flow. These flow rates were made from water levels attained in a discharge sump equipped with a V-notch weir. A float transducer converted water levels to current for recording. This type of flow measurement was found to be the most reliable and trouble-free of the several methods utilized. The four-inch, treated water return line to the plant water system also served as a sample line to monitor the final effluent pH, turbidity and conductivity.

In the rotary limestone reactor, horizontal lifters were included to aid desired stone attrition. They were not originally included in the fourth compartment to minimize wear to the exit port and screen. Although this purpose had been served, the small stone which accumulates in this section did not receive adequate attrition. The addition of four-inch wide lifters to this compartment improved reactor performance. Dams were also placed in the reactor to control stone distribution. This function was further aided by the addition of rod screens of three-inch, two-inch and one-inch openings across the apertures from influent to effluent ends of the reactor. The reactor slope provides some control on retention time and stone distribution. Other details are given in Appendix B.

To ensure minimal seepage from the raw water holding lagoon, bentonite clay was worked into the lagoon bottom and the surface compacted (see Appendix B, Figure 2).

In the design and operation of equipment to transfer air (thus oxygen) and subsequent evaluation of process data in iron II oxidation, the solubility of oxygen in water becomes important. Published data[11]

Table 9. CONSTRUCTION COSTS FOR HOLLYWOOD MINE DRAINAGE TREATMENT FACILITY
Dollars

	Equipment	Construction	Process Piping	Electrical	Heating	Ventilation	General Installation	Total Installed Cost
Pumping Station								
Bennett's Branch	3,987	17,961	1,925	4,000	250	200		28,323
Proctor No. 1	4,505	14,646	3,600	2,000	250	200		25,201
Proctor No. 2	5,276	19,190	3,710	3,000	250	200		31,626
Force Mains								
Bennett's Branch		3,850	11,600					15,450
Proctor No. 1			4,800					4,800
Proctor No. 2			5,600					5,600
Control Building	5,739	124,000	49,560	39,300	6,000	3,400		259,696
Laboratory	2,356							
Plumbing			29,341					
Service Building		18,150	1,500	2,800				22,450
Yard Equipment	16,196							
Densator	37,527	14,400	10,000	7,000				68,927
Flash Mixer	4,440						9,000	13,440
Holding Lagoon		12,311	1,760					14,071
$Ca(OH)_2$ Handling System	20,989	5,000		1,900			12,100	39,989
CaO Handling System	21,489	3,700		1,600			12,100	38,889
$CaCO_3$ Reactor	29,500	5,300	4,350	4,350				43,500
Oxidation Tanks	11,793	15,230	7,740	3,870				38,633
Surface Reactor	3,354	16,900	5,200	2,600				28,054
Settling Lagoon		30,220	8,586	4,293				43,099
Sludge Pumping Station	1,574	10,029	8,350	2,200				22,153
Sludge Drying Beds		17,829	3,000					20,829
Vacuum Filter System	41,142						36,700	77,842
Polymer System	2,212						6,400	8,612
Instrumentation	35,322			5,295				40,617
							TOTAL	891,801

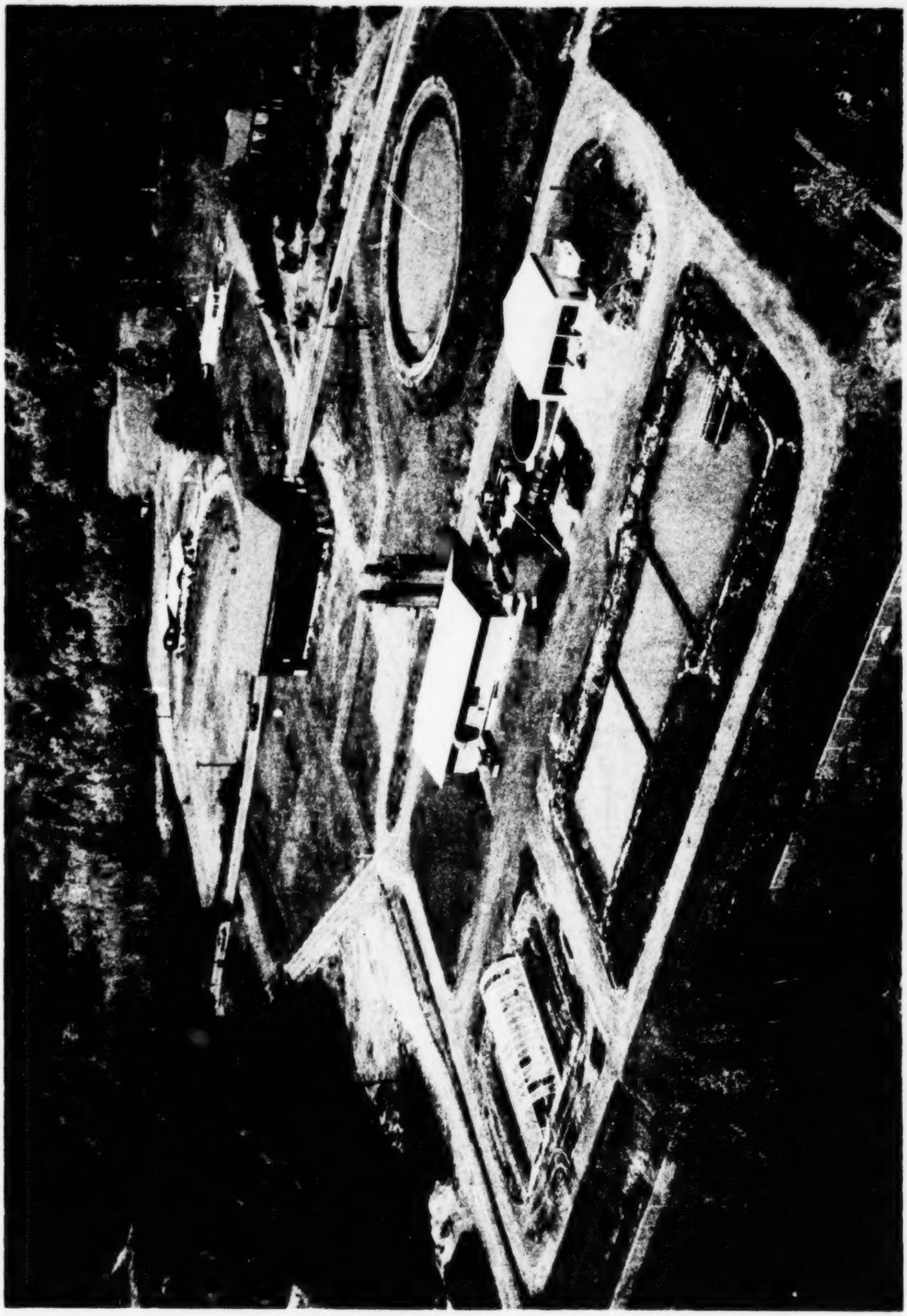


Figure 4. Aerial photograph of the experimental mine drainage treatment facility

Figure 5. Hydraulic profile of the experimental mine drainage treatment facility

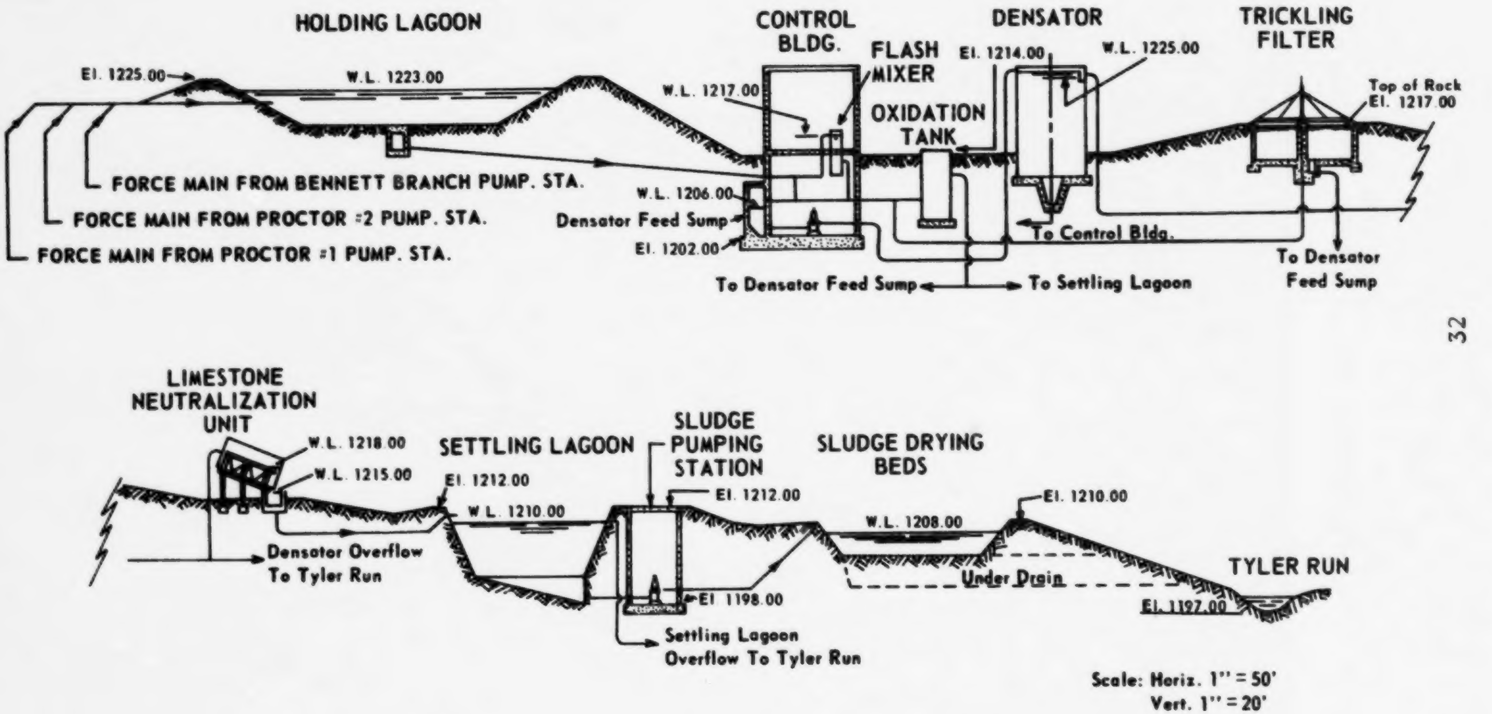
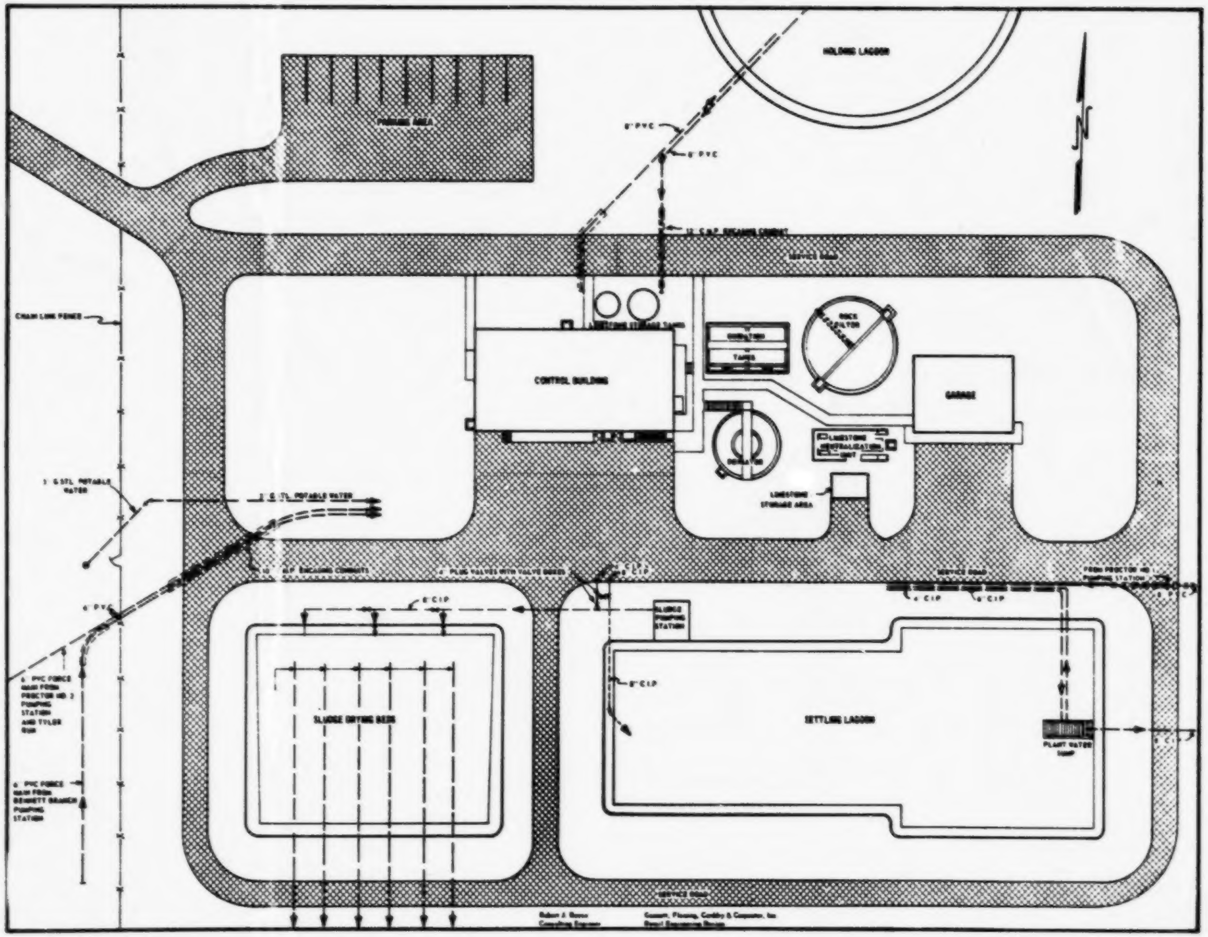
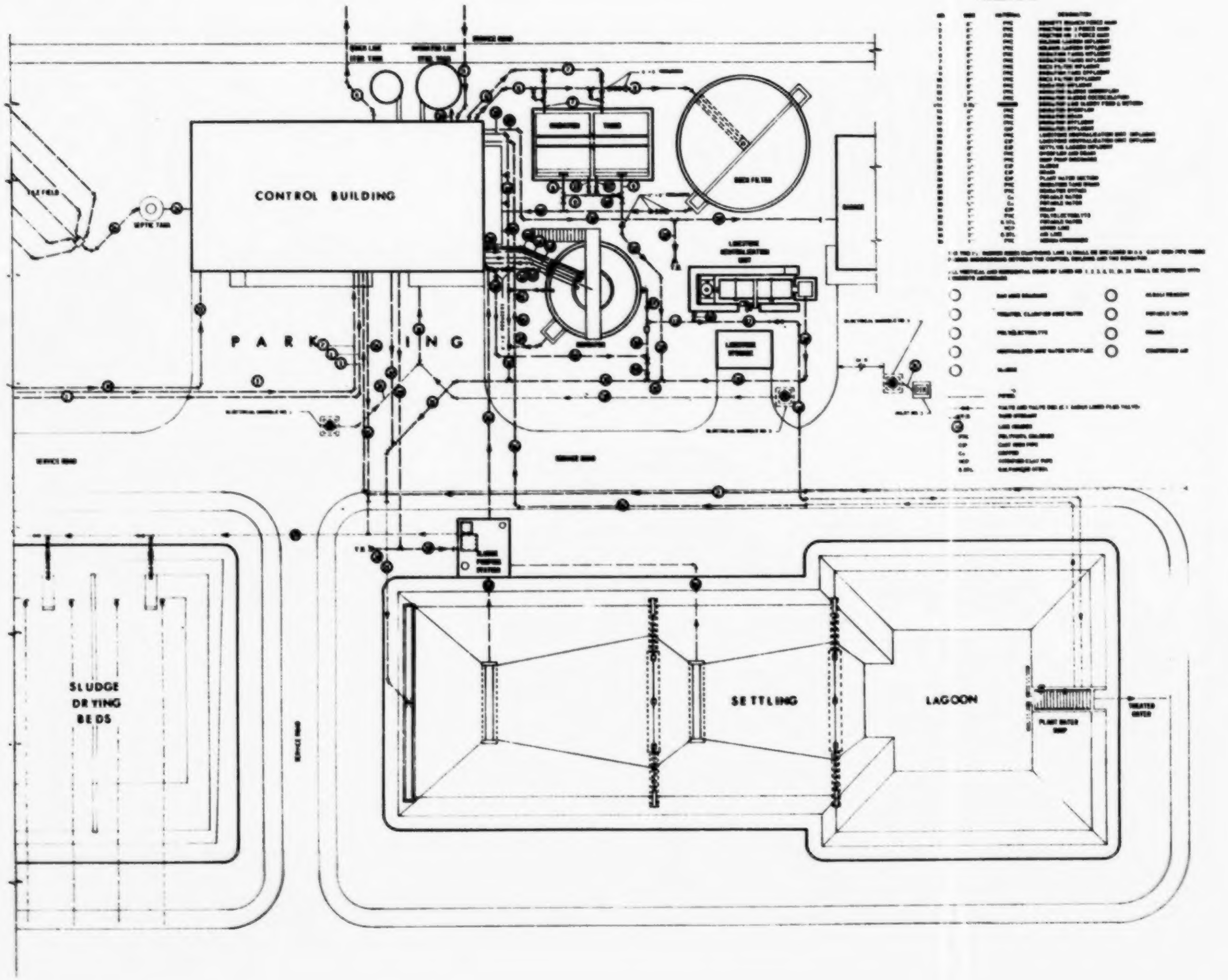

HOLDING LAGOON
El. 1225.00
W.L. 1223.00
FORCE MAIN FROM BENNETT BRANCH PUMP. STA.
FORCE MAIN FROM PROCTOR #2 PUMP. STA.
FORCE MAIN FROM PROCTOR #1 PUMP. STA.
CONTROL BLDG.
W.L. 1217.00
W.L. 1206.00
Densator Feed Sump
El. 1202.00
FLASH MIXER
OXIDATION TANK
DENSATOR
El. 1214.00
W.L. 1225.00
To Control Bldg.
To Densator Feed Sump
To Settling Lagoon
TRICKLING FILTER
Top of Rock El. 1217.00
To Densator Feed Sump
LIMESTONE NEUTRALIZATION UNIT
W.L. 1218.00
W.L. 1215.00
Densator Overflow To Tyler Run
SETTLING LAGOON
El. 1212.00
W.L. 1210.00
SLUDGE PUMPING STATION
El. 1212.00
El. 1198.00
Settling Lagoon Overflow To Tyler Run
SLUDGE DRYING BEDS
W.L. 1208.00
El. 1210.00
Under Drain
El. 1197.00
TYLER RUN
Scale: Horiz. 1" = 50'
Vert. 1" = 20'

Figure 6. External unit operations of the experimental mine drainage treatment facility



HOLDING LAGOON
N
PARKING AREA
6" PVC
6" PVC
12" C.M.P. ENCASING CONDUIT
SERVICE ROAD
CHAIN LINK FENCE
LIMESTONE STORAGE TANKS
CONTROL BUILDING
OXIDATION TANKS
ROCK FILTER
GARAGE
LIMESTONE NEUTRALIZATION UNIT
LIMESTONE STORAGE AREA
1" G.STL. POTABLE WATER
2" G.STL. POTABLE WATER
10" C.M.P. ENCASING CONDUIT
4" PLUG VALVES WITH VALVE BOXES
C.I.P.
SERVICE ROAD
FROM PROCTOR NO. 1 PUMPING STATION
4" C.I.P.
6" C.I.P.
6" PVC
8" C.I.P.
SLUDGE PUMPING STATION
6" PVC FORCE MAIN FROM PROCTOR NO. 2 PUMPING STATION AND TYLER RUN
SLUDGE DRYING BEDS
8" C.I.P.
SETTLING LAGOON
PLANT WATER SUMP
6" PVC FORCE MAIN FROM BENNETT BRANCH PUMPING STATION
SERVICE ROAD
Robert J. Rosso Consulting Engineer
Gannett, Fleming, Corddry & Carpenter, Inc.

Figure 7. External piping arrangement of the experimental mine drainage treatment facility


CONTROL BUILDING
P A R K I N G
SLUDGE
DRYING
BEDS
SETTLING
LAGOON
SEPTIC TANK

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not only indicates the variation in solubility with pressure and temperature, but that the oxygen-nitrogen ratio of air dissolved in water differs from that in the atmosphere.

More recent data for the solubility of oxygen in water in contact with air at one atmosphere pressure (air and water vapor) has been determined by Truesdale, et al.[12a]. The solubility, C, may be calculated with the following formula:

$$C(\text{ppm } O_2) = 14.161 - 0.3943t + 0.0077t^2 - 0.00006t^3$$

where t = °C

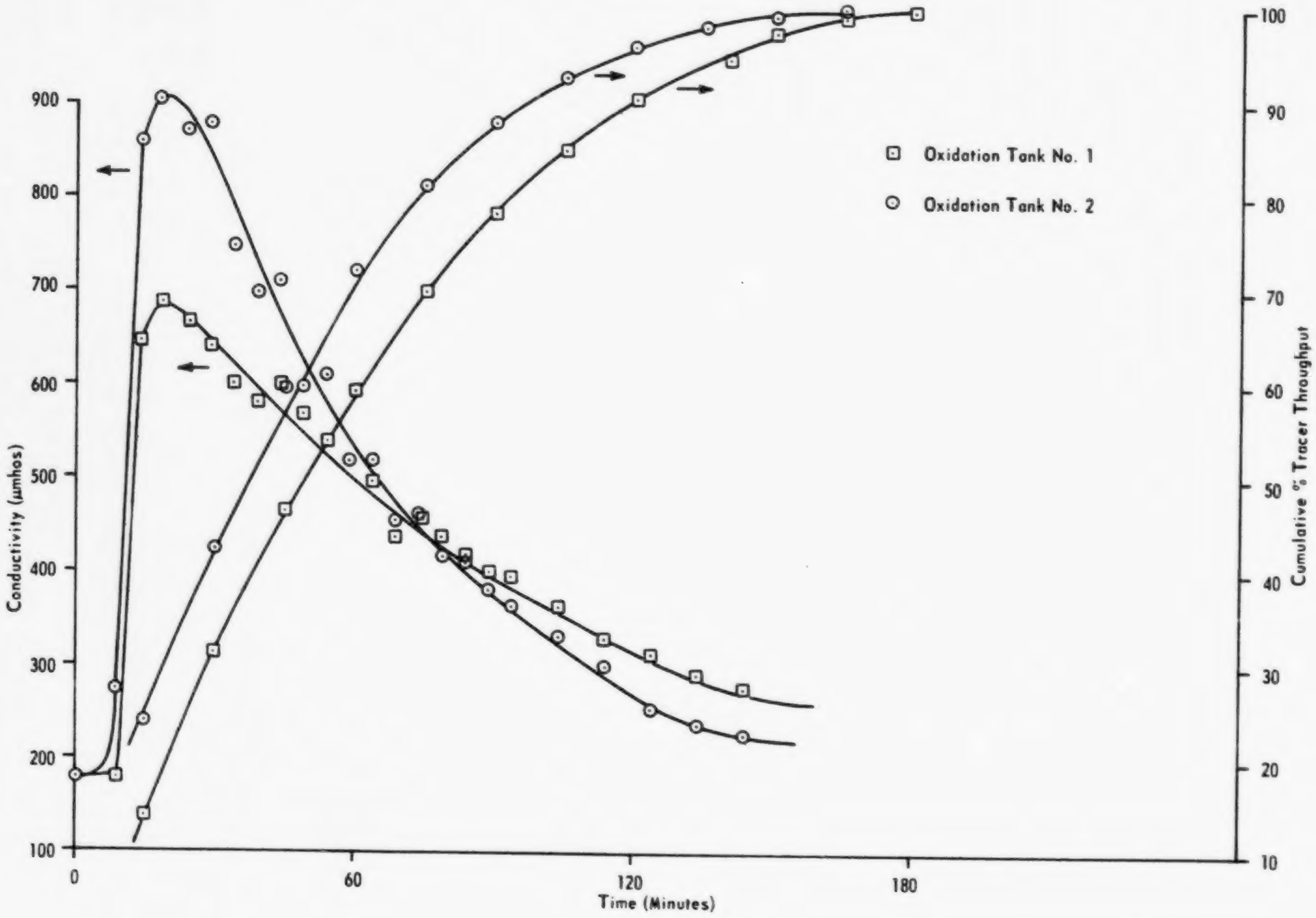
The literature indicates the solubility of oxygen in water is depressed by approximately 10 percent from soluble compounds (salts, acids, and bases) at concentrations of about 0.1N (5,000 to 15,000 ppm). At the salt and acid concentrations found in most coal mine drainage, the suppression probably would not exceed 5 percent of calculated values.

The flow response through and retention time in the oxidation tanks was evaluated at 200 gpm by a single rapid addition of sodium carbonate solution to Tyler Run waters in the flash mixer. The solution was directed to the oxidation tanks opened for parallel flow (see Figure 7). The oxidation tank effluents were sampled and conductivity measured as a function of time. The nominal retention time for the nearly 10,000 gallon vessels under these conditions was almost 50 minutes. Oxidation Tank No. 2 received the influent somewhat more rapidly and thus had a shorter retention time. The conductivity data are plotted in Figure 8 including an integrated plot representing tracer through-put. The first portion of tracer passed through Tank No. 2 in about eight minutes but required nearly thirteen minutes for the first tank. Half the tracer had passed the second tank in 38 minutes while this value for the first tank was 49 minutes. The highest concentration of tracer was attained at the effluent of each tank after about 18 minutes indicating this period to represent the retention time of the greatest number of tracer molecules. Over 99 percent of the tracer had passed Tank No. 2 in 140 minutes with 160 minutes being the corresponding value for the first tank.

The flash mixer, utilized to mix raw water with the slurried alkali reagent, was a rubber-lined steel vessel separated by baffles into three equal volume compartments as shown in Figure 9 and in Appendix B, Figure 9. The flash mixer had a minimum volume of 1,481 gallons.

The three-compartment sludge drying basin had overall dimensions of 103 ft. x 84 ft. Details of construction are shown in Figures 6, 7, and Appendix B, Figure 12. Photographs in Figures 4, and 46 through

Figure 8. Retention time studies - oxidation tanks



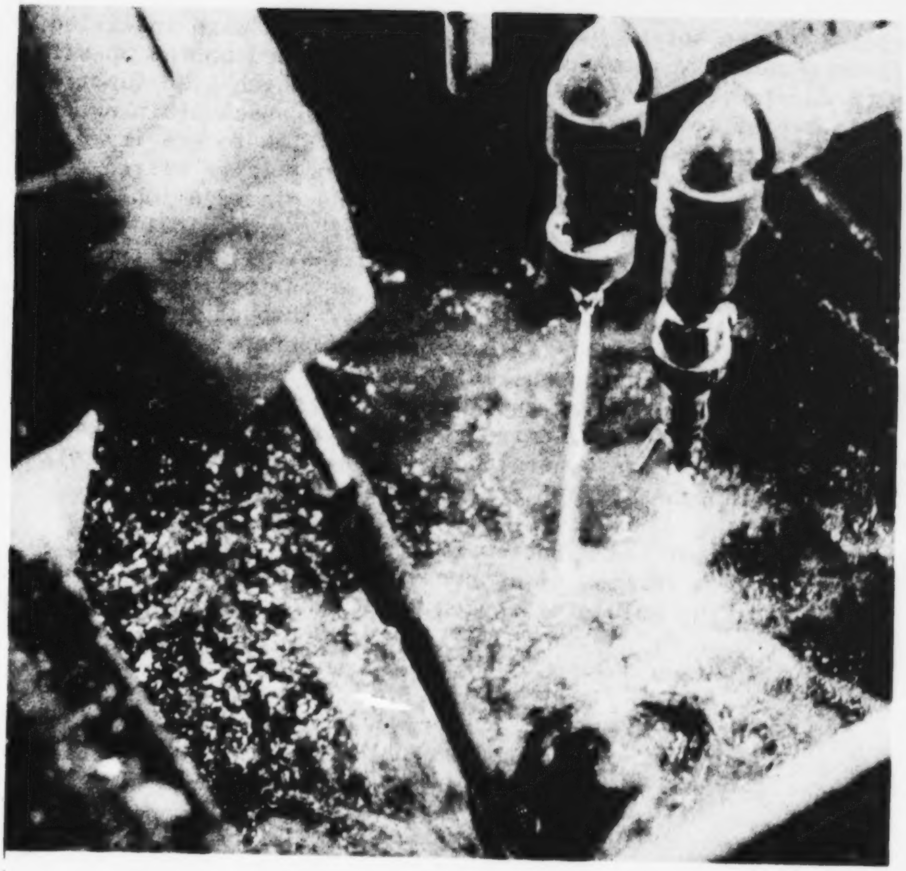


Figure 9. Introduction of Lime Slurry into Flash Mixer.

52 offer additional information. Basically, the bed was an earthen-walled excavation having three sides with a 30° slope. The south side, designed as an access ramp, had a 15° slope and was covered with four inches of crushed stone upon which was laid a two-inch bituminous surface.

The bed bottom was the prevailing clay base, carefully compacted. Each bottom section contained two 4-inch vitrified clay pipe drain lines of extra strength, bell and spigot-type with perforations restricted to the bottom half of the pipe. These were installed on 12 1/2-foot centers. Each drain line was equipped on the up-stream end with an air vent stand pipe six-feet high which also functioned as a decant pipe to remove excess clear water. Each drain extended through the south bed wall carrying the underflow to gravity drainage. The drainage pipes were covered with a two-foot layer of red dog (naturally-burned coal refuse) upon which was placed a four-inch layer of coarse sand. The drainage rates of the originally-constructed bed were excellent. During the heaviest rains, prior to the introduction of sludge into the bed, there was *no* collection of water in the drying bed with the water flowing freely through the underdrain. The bed was divided into equal sections by concrete walls designed to support a 6-mil polyethylene plastic canopy over the center section. The plastic was supported by an aluminum arch frame. The ends of the canopy were open to maintain free air circulation. The precipitation rejected from the central compartment by the plastic cover was collected in run-off gutters leading to natural drainage.

Each bed compartment had a bottom filtration area of 1,270 ft^2 and could accomodate a maximum of a three foot depth of sludge, thus providing a working volume of approximately 4,700 ft^3 (35,000 gallons).

SECTION V

OBJECTIVES

This study resulted from a request of the Pennsylvania Coal Research Board to design, construct and operate a full scale plant to evaluate treatment procedures for coal mine drainage waters. The study was to provide year-round data covering operating and economic experience. The Hollywood Facility was developed to provide maximum versatility for the study of neutralization treatment processes and related procedures. Special emphasis was placed upon control and dewatering of the sludge which represents the impurities removed from the contaminated waters.

This site permits the study of four waters of different qualities as a process parameter. All basic stages proposed to treat these waters by neutralization can be reproduced. In addition to the defined program, the Facility provides the opportunity for detailed study and observation of any specific unit operation. The available basic systems include:

1. The "Yellowboy" Process A: Neutralization by an alkaline reagent (hydrated lime, calcined lime, caustic soda, soda ash, limestone, dolomite products, and waste lime products) in a flash mixer, air oxidation of ferrous iron, and separation of the sludge by the settling lagoon.

2. The "Yellowboy" Process B: Neutralization by an alkaline reagent, air oxidation of ferrous iron, and separation of the sludge by clarifier.

3. Infilco Densator Sludge Recycle Process: The entire treatment proceeds in a specially designed reactor in which the raw drainage water is mixed with recycle sludge in the primary compartment, alkali added and reacted in the secondary compartment, the sludge separated in the bottom and subject to recycle or removal. Finally the treated, clarified water is discharged from the top.

4. Biochemical Oxidation-Limestone Process A: Deep tank oxidation of iron in the raw water, separation-recycle-removal of biologically-active sludge in the clarifier, water neutralization with limestone, and separation of sludge in the settling lagoon.

5. Biochemical Oxidation-Limestone Process B: Surface oxidation of iron in the raw water, separation-recycle-removal of biologically-active sludge in the clarifier, water neutralization with limestone, and separation of sludge in the settling lagoon.

6. Limestone Neutralization: Direct chemical oxidation and neutralization in the limestone reactor and separation of sludge in the settling lagoon.

SECTION VI

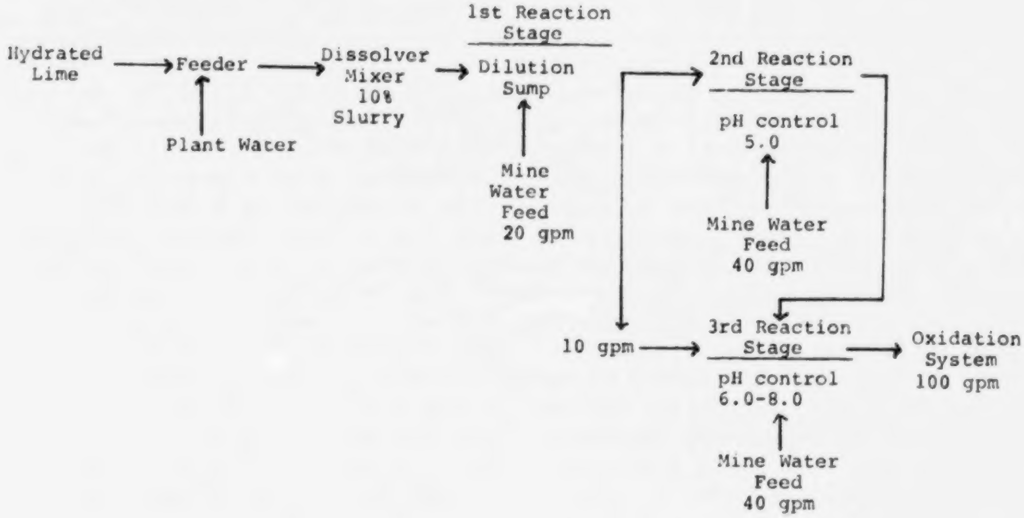
TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME

The process most commonly utilized to treat coal mine drainage mixes the water with calcium hydroxide, raises the pH and forms insoluble iron and aluminum compounds. The reaction mixture is subsequently oxygenated converting iron II to iron III. The suspended solids are then separated, discharging the treated water.

The studies, considering each of the four waters available, utilized this procedure with two modifications of solids separation: settling lagoon (Process 10) and thickener (Process 11). The process flowsheet and test data for Process 10 are detailed in Appendix C, Tables 1 through 4. In these tests the controlling pH probe was located in oxidation tank No. 2. The data for Process 11, where the pH probe was located in the flash mixer, are given in Appendix C, Tables 5 through 9. The reagent was introduced as an aqueous slurry containing between one and five percent calcium hydroxide as shown in Figure 9. The system permitted the lime slurry to be controlled and introduced either by a manual valve or by a proportioning valve which responded to a setting of the pH probe. See Appendix B, Figure 9. A constant manually-controlled slurry addition was utilized when treating large volumes of the more highly mineralized water with the concurrent addition of lime slurry through the proportioning automatic valve. This procedure yielded minimal pH variation within the system. The manual valve alone did not provide reliable, constant slurry flow. The lime slurry feed system was effective in providing a continuous, controlled flow completely automatic from lime storage bin. No difficulties were experienced with lime scaling in the lines or hang-up in the bins. Due to intermittant operation and prevailing high humidity, the electronic controls of the feeder required frequent servicing. Some carbonation of the lime slurry occurred in the lime slurry sump, especially in those operations where the lime consumption was low and the slurry was prepared at infrequent intervals. This introduced an uncertainty in reagent measurements, especially with short runs. The test program necessitated that some batches of hydrated lime remain in the storage bin for weeks. Although no handling or moisture absorption difficulties developed, there were major levels of reagent carbonation which changed the character of the alkali reagent, and its reaction mechanism. In operating treatment facilities, the lime storage capacity should not exceed one month's requirements.

The addition of lime (or any of the reagents) as an aqueous slurry avoids delays in wetting the solid. The slurry mixes with the water more rapidly. The solubilized lime reacts on contact while those solid particles remaining are more rapidly dispersed for solution and reaction. It appears desirable to add the reagent at the most dilute level feasible. The total reagent requirements establish the limits. This rapid mixing and reaction permits a more responsive action of the controlling pH probe. Localized excesses of reagent are less likely to occur thus avoiding unwanted side reactions. Reagent addition control is most versatile when a slurry is employed. Large clumps of unwet lime are more likely to exist with dry reagent feeding and to remain together leading to possible coating and to loss of unreacted reagent. Addition of solid reagents was not utilized in these studies.

The lime slurry was prepared with either house water or treated effluent. The preparation of a concentrated (10-20 percent) reagent slurry with good quality water (as in the mixer-dissolver), to be diluted with the water being treated, is indicated. The dilution would become the first of several partial neutralization stages. In the Hollywood design this could be accomplished in the lime slurry sump. The level of dilution would vary with the acidity of the feed water but the final slurry probably should never exceed 0.1 percent by weight. This procedure would offer the best control to maintain the desired uniform pH and avoid excess alkali consumption by unwanted reactions. An even more desirable reaction situation would follow if the diluted reagent were added to the water in several stages as follows:



Hydrated Lime
Feeder
Plant Water
Dissolver Mixer 10% Slurry
1st Reaction Stage
Dilution Sump
Mine Water Feed 20 gpm
2nd Reaction Stage
pH control 5.0
Mine Water Feed 40 gpm
10 gpm
3rd Reaction Stage
pH control 6.0-8.0
Mine Water Feed 40 gpm
Oxidation System 100 gpm

A theoretical explanation for this recommendation, which was based on these laboratory and plant observations, is offered by Whittemore[70]. This mode of reaction also affects sludge behavior.

In the Process 10 tests, the pH control was maintained at the lowest pH possible, consistent with meeting final iron specifications. In some cases the acidity requirements were not maintained unless significant reagent excesses were utilized. In those cases (as Test No. 8-70), the alkalinity developed in the effluent was less than that indicated by the excess reagent. This suggests that other reactions, probably involving aluminum and magnesium, were consuming reagent. This observation is important in considering reagent consumption in respect to theoretical requirements and treatment objectives. The goal to reduce iron concentration and develop some alkalinity involves aluminum and magnesium concentration secondarily. The reaction of further lime with aluminum hydroxide beyone its minimum solubility level (about pH 6.2) appears to lead to its resolubilization as the aluminate. This means less removal of aluminum from the waters and higher lime consumption. Despite theoretical considerations, magnesium does appear to react at unexpected low pH levels, possibly precipitating and reducing magnesium concentration in the effluent. This consumes lime beyond the level related to established goals. The desirability of these effects requires a management decision: use of minimum reagent (reduced costs) versus less aluminum and more magnesium removal.

This situation was more evident in Process 11 tests, where the pH probe was located in the flash mixer. Two probe locations in this reactor were utilized. The pH variation was found to be greater (more excess reagent utilized) when the probe was located near the flash mixer overflow (Tests 66-71, 67-71, 68-71, 69-71 and 71-71), as opposed to a location in the same compartment where the alkali was added (Tests 38-71, 56-71, 57-71 and 65-71). The excess of reagent consumption tends to increase with the acidity of the raw water and with smaller flows where reaction time was longer.

The Process 10 tests generally produced satisfactory effluent meeting the specifications: pH > 6.0, slight excess alkalinity (several tests resulted in low acidity values), iron < 4 mg/l, and a clear effluent with suspended solids < 12 mg/l. The reagent consumption decreased in efficiency with increasing acidity, ranging from 54 to 192 percent. The adjustment to levels near theoretical was feasible but was difficult during these short duration runs. The sludge production also varied with the total acidity of the feed water. The settled sludge (two hour level) ranged from 0.14 to 0.52 weight/volume percent, the dry solids from 0.23 to 2.84 lbs/1000 gallons and the sludge volume from 1.2 to 14.5 volume percent. The sludges displayed Type I settling rate behavior (see Section XIV) which tended to gel upon aging. Settling rate data are given in Figures 24 and 25, while summary sludge settling characteristics are given in Table 24. The response to separation in the settling lagoon is provided in Tables 39 to 42, while the data from a thickener are detailed in Tables 34 and 35. Chemical analyses of the sludges are listed in Table 50. The iron II oxidation was satisfactory at the prevailing retention times, ranging from 0.56 to 30 minutes/ppm iron II.

The Process 11 tests were also satisfactory, with more emphasis on higher acidity waters, but at lower flow rates due to the thickener size. The effluent pH was maintained at higher levels than the Process 10 (settling lagoon) tests, always exceeding 7.0. The alkalinity values ranged to 345 but were generally near 14 mg/l. The soluble iron was consistently less than 1 mg/l. The effluent was sparkling clear without the use of flocculants. The reagent consumption ranged to 283 percent of theoretical with control becoming more difficult with the higher acidity waters. Adequate adjustment was obviously feasible. In Proctor No. 2 waters the sulfate content was reduced by 22 percent, as was magnesium by nearly 50 percent. The two hour sludge content was similar to the previous tests (0.42 to 0.69 weight/volume percent) but increased to 2.6 percent after 10 hours. The dry solids were higher with Proctor No. 2 waters ranging to 14.2 lbs/1000 gallons treated water as was sludge volume extending to 26 volume percent. The oxidation retention time was increased but proved to be adequate at levels as low as 0.37 minutes/ppm iron II. The oxidation-reduction potential ranged between 162 and 350 millivolts. The minimum oxidation rate extended to 338 mg/l/hr in these runs.

In contrasting the two unit operations for solid-fluid separation (settling lagoon versus thickener), as applied with this reagent, there are no obvious differences which affect the chemical treatment. The plant experiences from these solid-fluid operations and subsequent dewatering are presented in Section XIV.

SECTION VII

TREATMENT OF COAL MINE DRAINAGE WITH CALCINED LIME

Tests with calcined lime followed the flowsheet (termed Process 10--see Table 10, Appendix C) used with hydrated lime (Tables 1 through 4, Appendix C) with the replacement of a slaker for the dissolver-mixer. Each of the four waters was treated in the flow range from 101 to 325 gpm. The slaked reagent was diluted to the one to five percent level for pumping to the flash mixer. The average percent concentration of $Ca(OH)_2$ for runs 6-71 through 11-71 were, respectively, 2.8, 0.93, 1.70, 2.30, 1.31 and 1.18. The slurry addition was controlled as described with hydrated lime.

The pH probe controlling slurry addition was moved from the oxidation tank to the first compartment of the flash mixer after Test 7-71. For Test 15-71 it was moved to compartment No. 3 near the overflow pipe. As with the hydrated lime tests, proximity of the control probe to the slurry feed point tended to result in greater alkali consumption. The difficulty of pH control (as evidenced by range of pH observed) tended to increase as retention time decreased as shown in Table 10. These values are extremes and do not reflect the stable control attained which was normally less than ±0.5 pH units. These variations with calcined lime were greater than hydrated lime, reflecting a slower level of reactivity. The feed water was 8 to 17°F colder during these studies than the hydrated lime tests which would also decrease reactivity. Otherwise there were not any distinct operational differences observed. Calcined lime is the most commonly used reagent other than hydrated lime.

The storage, conveying and slaking of calcined lime gave a satisfactory response without mechanical difficulties. However, it required additional attention to remove the unreactable grit and to maintain slaker operation. Consideration of known variations in slaking characteristics with different calcined limes and attainment of optimum slaking conditions was beyond the scope of this study. The non-slaking grit represents unburned limestone. The four to eight percent loss of grit, its handling, equipment maintenance as well as its capital costs minimize the cost advantage of this reagent. The grit products not only vary with the calcined lime product but with the operation of the slaker. In test 12-71 the grit production was 9.2 percent while in test 13-71 it was 5.2 percent. The slaker must be kept at hydration reaction temperatures (>130°F); thus it lacks the versatility of intermittant operation feasible with the hydrated lime dissolver. The slaking temperature ranged between 102 and 210°F with an average of 143°F. Failure to maintain a proper slaking temperature tends to result in loss of alkalinity

Table 10. SUMMARY OF pH CONTROL VARIATION
CALCINED LIME TREATMENT - PROCESS 10

Test no.	Oxidation tank response time minutes	Flow rate GPM	pH Range Flash mixer	pH Range Oxidation tank
17-71	<5	325	8.0	7.7
7-71	10	197	6.8	1.9
10-71	minimal	196	1.6	2.5
12-71	10-15	199	1.4	2.3
15-71	15	200	1.9	2.3
14-71	15	199	4.0	4.1
13-71	15-20	195	1.0	1.2
9-71	20	107	3.2	2.8
8-71	<30	109	7.2	6.1
11-71	60	103	1.9	0.9
16-71	60	101	4.0	2.5

with the grit and a less reactive reagent. The lime consumption must be at least adequate to maintain continuous operation of the smallest slaker commercially available (about 300 lbs/hr and a production of 12 to 25 lb/hr of grit). Only in a few tests was this rate approached. Calcined lime is favored cost-wise as the reagent requirements of the system increase.

The detailed test data are given in Appendix C, Tables 10 through 14. The observations during the hydrated lime tests were confirmed regarding advantages of stepwise addition of the alkali to prevent localized overtreatment and subsequent reactions with aluminum and magnesium. These responses tend to consume excess reagent. There was no indication of moisture or carbon dioxide absorption by the calcined lime during storage. Calcined lime must be utilized as a slurry to achieve an adequate utilization and reaction rate. A recently developed, pulverized, calcined lime product is claimed to react with sufficient rapidity that separate slaking is not necessary. This special product was not available for these tests.

During these runs the pH control was adjusted to maintain a level in excess of 7.0 to avoid any residual acidity. This control, believed to be typical of industrial practice, leads to considerable excess reagent consumption. The utilized excess increases with raw water acidity and tends to increase at lower flow rates with longer reaction times, ranging from 77 to 272 percent. These excesses were not detectable as alkalinity in the effluents. Those tests showing very high reagent consumption used a raw water having relatively large concentrations of aluminum and magnesium.

Despite the relatively high pH levels maintained and the excess reagent utilized, the settling lagoon effluents tended to develop small levels of acidity. The iron content was characteristically 1 mg/l or less. The effluents were sparkling clear with suspended solids seldom exceeding 10 mg/l.

As with the hydrated lime tests, the sludge production increased with total acidity of the feed water. The solids level of two hour settled sludge was somewhat higher (0.20 to 0.95 weight/volume percent) than with hydrated lime, as was dried solids production (0.55 to 19.1 lbs/1000 gallons). The settled sludge volume was consistent with the hydrated lime tests, varying between 1.3 and 29.6 volume percent. The settling behavior of the sludges was typical Type I excepting those developed from Tyler Run and Bennett's Branch waters which responded as Type II. These sludges tend to gel upon aging. See Figures 24 and 25 for typical settling curves and Table 24 for sludge settling data. Tables 39 through 42 show the response of these sludges in the settling lagoon.

The iron II oxidation was adequate at the retention times employed (0.19 to 196 minutes/ppm iron II). The largest minimum oxidation rate observed was 416 mg/l/hr. The measured oxidation-reduction potentials were somewhat higher (increased pH levels) than observed with hydrated lime extending from 198 to 414 millivolts.

To summarize the responses shown by hydrated lime and calcined lime: there are no obvious deviations. There is the very critical cost difference between them but this appears advantageous only when the reagent consumption is large enough to accomodate the additional equipment and maintenance necessary to slake the calcined lime.

SECTION VIII

TREATMENT OF COAL MINE DRAINAGE TO PRODUCE HIGH DENSITY SLUDGE BY RECYCLE

The complications resulting from the voluminous, low solids content sludges which result when most alkalies are reacted with coal mine drainage is cited throughout this study. The difficulties became obvious when seeking daily disposal of 300,000 gallons of a slurry containing less than one percent suspended solids resulting per million gallons of coal mine drainage. The observed results from secondary dewatering attained with the sludge drying basin (Section XIV) indicate this approach to be feasible. The costs of some dewatering operations can be exorbitant. However, this method as well as other dewatering procedures, as multiple stage thickening, filtration and centrifugation, result in better efficiencies and lower costs if the density of their feed can be increased. Some improvement in settled sludge density can result from reagent selection (especially limestone) or reaction procedures (pH control and stage reagent addition). Polyelectrolytes were ineffective in increasing sludge density, although at high dosages a "balling" effect was observed which is worthy of further exploration. Seeking means to maximize sludge density was a goal of this study.

The concept of sludge recycle has been applied successfully to sewage sludges and was reported helpful when treating waste pickle liquors[13, 14]. The Infilco "High-Density Solids-Contact Treatment Process" was investigated and incorporated into the Hollywood Facility design. Preliminary bench scale studies were made by project personnel and at Infilco's Research Laboratories, utilizing mine water from the Hollywood sources. A three-foot diameter Densator pilot unit was operated at the Hollywood site and observed by industrial and governmental representatives. Subsequently, a similar full-scale system has been constructed and operated[15]. That operation confirmed these preliminary studies attaining high sludge densities. A number of operating parameters were cited.

The physical unit installed at Hollywood was a modified Infilco "Densator", High Density Solids Contact Treating Plant. Bottom rakes were added to permit the operation of the unit as a conventional clarifier-thickener, as described in Section XIV. A detailed drawing of the unit is shown in Appendix B, Figure 14 with photographs in Figures 4 and 5 (Appendix B). The system provided for controlled pumping of sludge from the bottom well to the Densator feed weir, to the sludge storage sump for subsequent dewatering, or to the sludge drying basin. The recycle sludge rate could be monitored by a magnetic flow meter. These flows may be followed in Figure 7 and Appendix B, Figure 9. Alkali slurry could be pumped to the secondary reaction zone of the Densator in response to a pH sensor. This probe sampled water from the secondary reaction zone. It was also possible to add flocculant to this zone of the reactor. The system permitted premixing either recycle sludge and raw water or recycle sludge and alkali before further reaction. The above ground installation was convenient for regional sampling of the Densator.

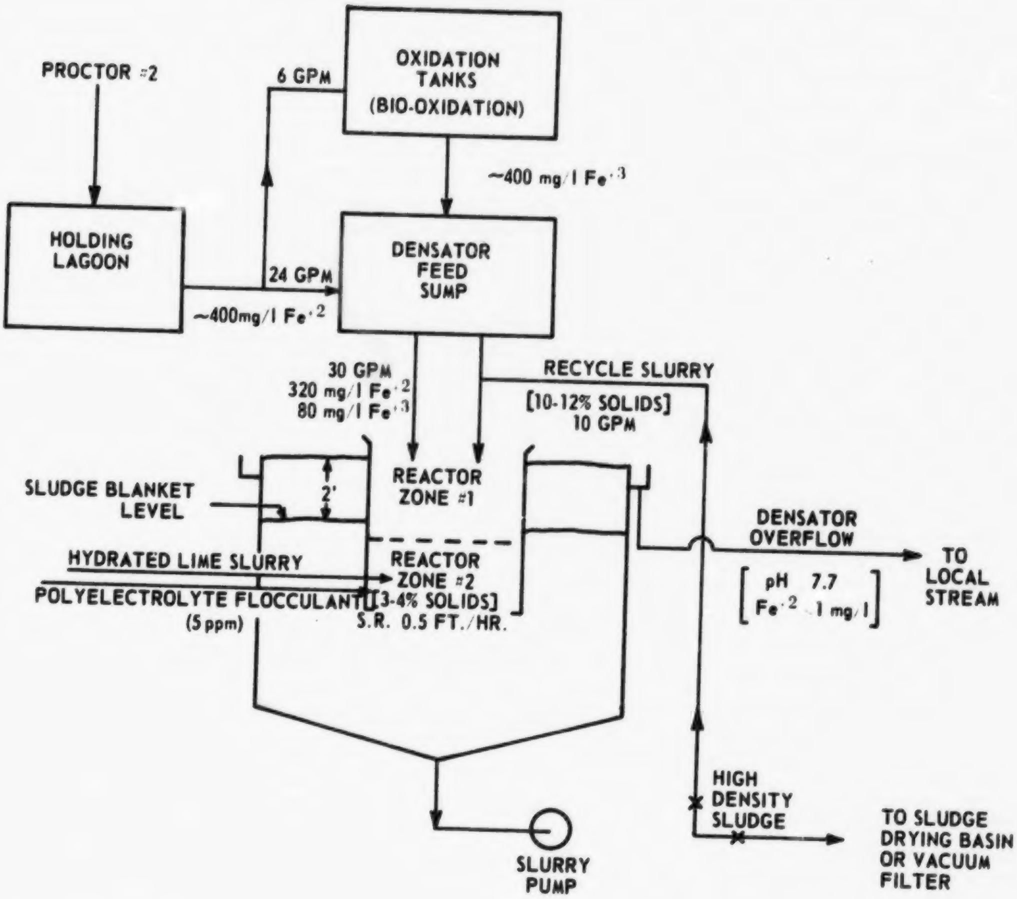
The single unit was designed to provide neutralization, sludge densification and clarification for a complete treatment facility. The sludge densification capability was most important. The test sought to demonstrate the potential of the Densator as a multifunction unit producing a densified sludge with good dewatering characteristics.

DENSATOR OPERATING CONDITIONS

The full-sized densification unit was not amenable to short time evaluation. The test described covered about one-hundred operating days necessitating a modification of the planned project program. Unfortunately, plant layout precluded simultaneous evaluation of other processes while the Densator was in operation, although some experience with biochemical iron II oxidation was carried out concurrently. The biochemical effluent, as indicated in the flowsheet (Figure 10), formed a portion of the Densator feed water. This concurrent study undesirably decreased the iron II/iron III ratio, complicating the analysis and control. Accordingly, the system was operated only with Proctor No. 2 water which was the most severely mineralized. The experience gained, including the laboratory and pilot tests, indicated the sludge recycle concept holds its greatest advantage when applied to highly mineralized drainage waters which characteristically form large volumes of low density (0.2-1.0 percent) settled sludges. Lightly mineralized waters, forming slow settling floc of negligible solids content, are less amenable to the process. Such waters tend to have low iron and aluminum content and normally produce less than one pound dry solids per 1000 gallons.

The Densator operation was controlled to maintain effluent quality consistent with Pennsylvania's regulations. This was accomplished. The sludge recycle rate sought to maintain approximately 3.0 percent solids in the primary reaction zone of the Densator. These solids were composed of a mixture of freshly-precipitated and recycled sludge. The Proctor No. 2 waters alone yielded about 1500 ppm suspended solids upon neutralization. To maintain the control

Figure 10. Flow sheet for producing high density sludge


PROCTOR #2
6 GPM
OXIDATION TANKS (BIO-OXIDATION)
~400 mg/l Fe+3
HOLDING LAGOON
24 GPM
~400mg/l Fe+2
DENSATOR FEED SUMP
30 GPM
320 mg/l Fe+2
80 mg/l Fe+3
RECYCLE SLURRY
[10-12% SOLIDS]
10 GPM
SLUDGE BLANKET LEVEL
2'
REACTOR ZONE #1
HYDRATED LIME SLURRY
POLYELECTROLYTE FLOCCULANT
(5 ppm)
REACTOR ZONE #2
[3-4% SOLIDS]
S.R. 0.5 FT./HR.
DENSATOR OVERFLOW
[pH 7.7
Fe+2 < 1 mg/l]
TO LOCAL STREAM
HIGH DENSITY SLUDGE
SLURRY PUMP
TO SLUDGE DRYING BASIN OR VACUUM FILTER

level of solids in the primary reaction zone, as was recommended by Infilco, an optimum ratio of recycled to precipitated solids was about 20:1. The feed water flow rate to the unit was established in conjunction with the other variables, primarily by the settling rate of the sludge blanket. The high density sludge concept functions from a highly ferrous system and alkali addition to maintain a pH level higher than conventional neutralization processes. A 4:1 iron II/iron III ratio was maintained during most of the test, although this ratio and total iron levels were decreased by rain dilution near the end of the test. A decrease in solid content of the underflow resulted from these concentration changes. No provision is made for iron oxidation in the system. This represents a capital and operating cost savings. Air introduced into the secondary reaction compartment interferes with sludge flocculation and settling rate. The monthly analyses for Proctor No. 2 waters during this test period are shown in Appendix A.

Hydrated lime slurry was pumped into the secondary reaction zone in the Densator to maintain a final pH of about 7.7. At this pH level, the iron concentration in the overflow was less than 1 mg/l. The values for the operating parameters are presented in Figure 11. A flow rate to the Densator in the range of 50 to 130 gpm permitted proper continuous operation, with the lower rates offering better overall response. The flow rate was modulated with the settling characteristics of the prevailing sludge settling rate. The use of a weakly anionic flocculant was desirable, if not necessary. The dosage was in the range of 4-5 ppm based upon influent volume. It appears that the flocculant dosages should be determined from the solids loading in the secondary reaction zone and expressed in terms of pounds of flocculant per ton of dry solids rather than on a raw water volume basis. Under these conditions, high density sludges ranging from 10 to 16 weight percent solids were produced consistently. The initial increase in sludge density was quite slow. The initially-formed sludges settled with typically low solids contents. Accordingly large volumes of settled sludge were recycled seeking to attain the desired three percent solids level in the reaction sections. Over 50 operating days were expended to meet the three percent criteria. As sludge density builds in the unit during this period, the normal sludge free settling rate is minimized by entering early stages of compression. It appears start-up should utilize low raw water and recycle flow rates, ignoring solids levels in the reaction sections. This would provide minimum rise velocities and permit greater sludge compression to develop in the sludge collection well.

No sludge density improvement was indicated by premixing the lime slurry with the recycle sludge before reaction with the raw water.

Figure 11. Operational data from densator recycle studies

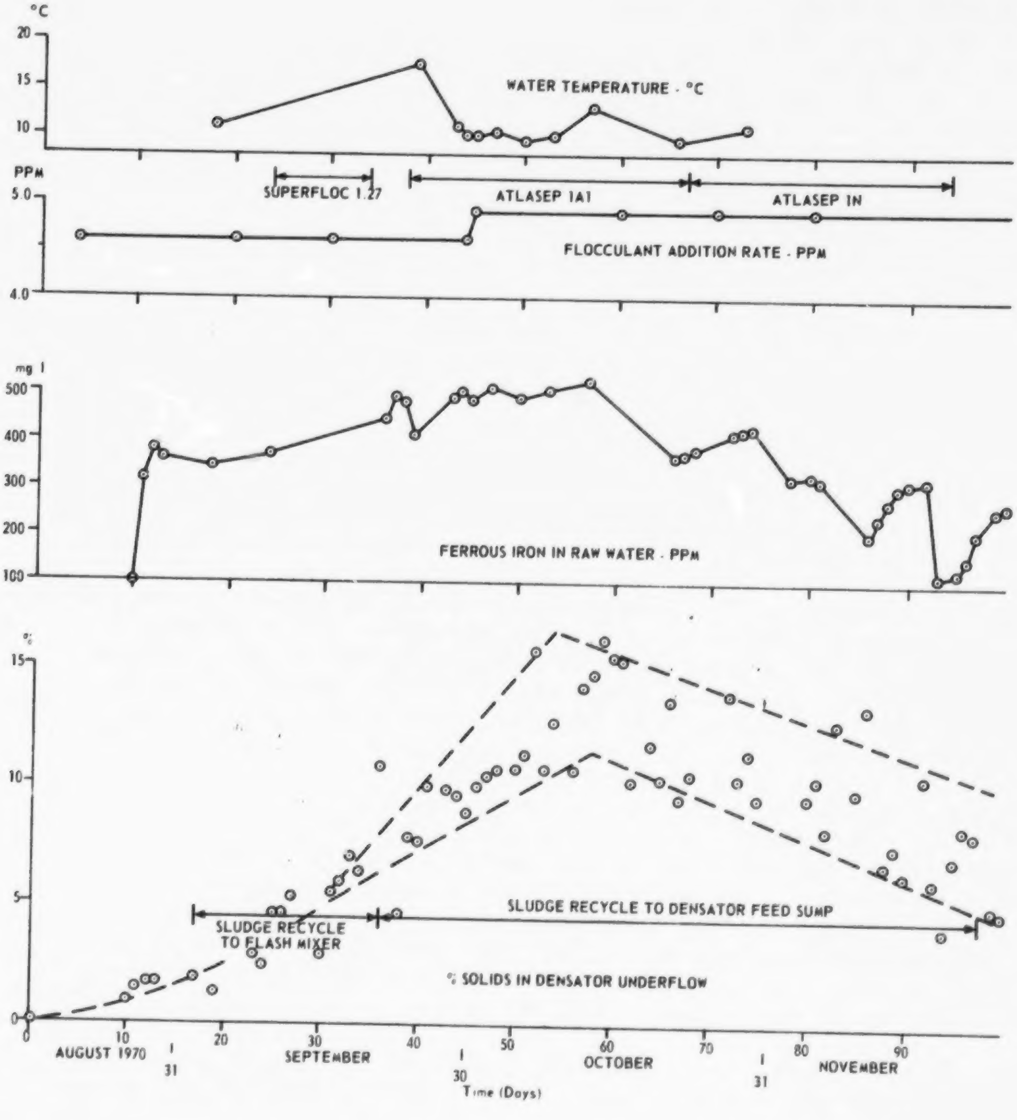
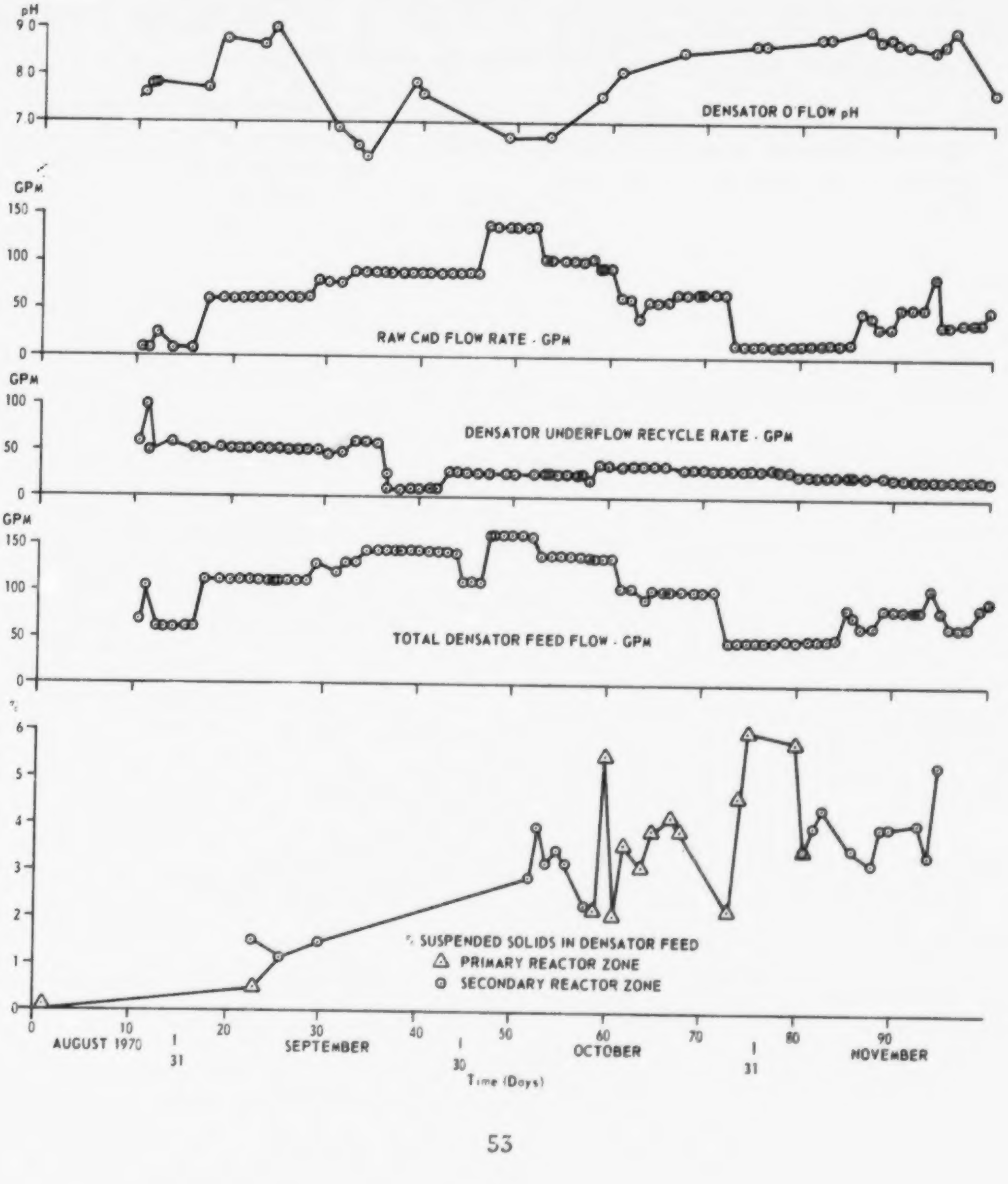


Figure 11 (continued). Operational data from densator recycle studies



Individual underflow samples showed solids content to nearly the thirty percent level. The samples were taken from a pumped stream. The solids content values cited represent recycle sludge and not that from additional settling. Recycle sludge densities varied widely during adjacent samplings. This sampling difference was associated with the position of the bottom rake, the depth of the sludge compression zone, and sludge flowability. The recycle sludge showed no tendency to rat-hole, as do gelled hydroxide sludges. Its viscosity reduces flow rates. A photograph of the recycle sludge (Figure 12) indicates viscosity in supporting a stirring rod upright. These sludges were ferrous in character (dark green).

No direct evaluation was made of the pollution potential of these ferrous recycle sludges after blow down. They are highly alkaline in character and oxidized very rapidly at the sludge-air interface when transferred to the sludge drying basin. This highly alkaline nature should be adequate to neutralize any acid formed during oxidation of the iron II content.

Water effluent quality was excellent during most of the test except when the control pH was allowed to drop below 7.5. Even under these conditions the overflow maintained an iron content consistently less than one mg/l.

A summary of the operating parameters is given in Table 11. A wide fluctuation in operating variables was experienced as the parameters were modified during preliminary operation of the system. It was difficult to maintain steady-state flow conditions because of the slow settling rates of the high level sludge blanket in the Densator. This sludge, which was well into the compression zone range, had an average settling rate of less than 0.5 ft/hr, which made it difficult to operate at continuous water flow rates much greater than 50 gpm. During these operations the level of the sludge blanket varied up to about 10 inches of the overflow. The input flow rate was the variable utilized to control the sludge blanket level. It was possible, however, to determine optimum conditions for the 25-foot diameter device. Based upon plant observations, optimum operating flow rates for the Hollywood Densator under various recycle conditions have been calculated and are presented graphically in Figure 13. These calculations are based upon a maximum sludge blanket settling rate of 0.5 ft/hr. The flow capacity of the Densator possibly can be increased by using greater flocculant dosage.

The recycle sludge produced during the tests had the desired properties sought. Besides high density (thus significant sludge volume reduction), the sludge had near ideal dewatering characteristics. Upon transfer to the sludge drying basin, the recycle sludge drained immediately, and never maintained any free standing supernatant. It

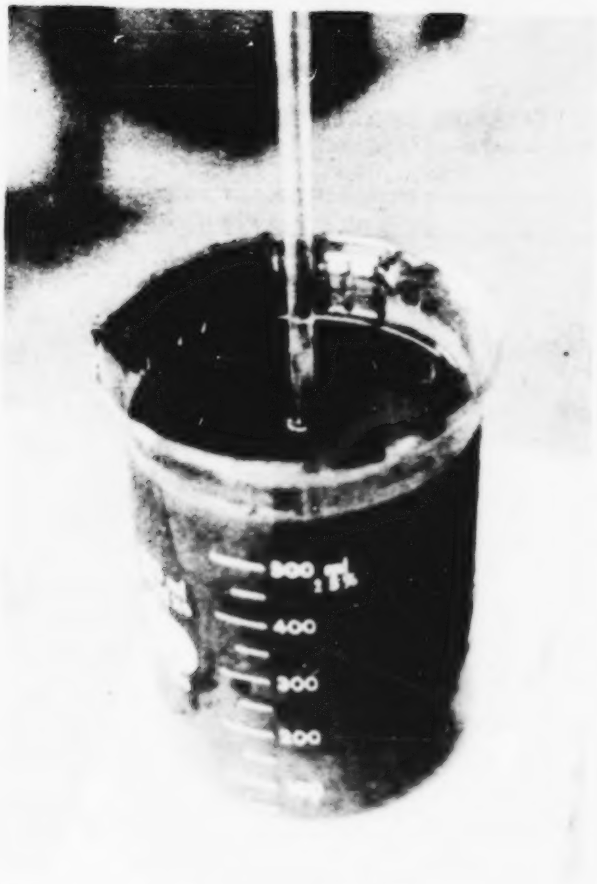


Figure 12. Mine Drainage Recycle Sludge.

also provided the best filtration characteristics of any sludge tested. With a filter feed suspended solids level between 10 and 17 percent solids by weight, a filter cake in excess of 1/2-inch could be maintained without blinding or build-up of filter vacuum. The cake cut cleanly from the drum with the very minimum of precoat. The filter cake had about 24.5 percent solids with excellent handling characteristics. The solids filtration rate averaged about 312 lbs dry solids/ft^2/24 hrs. This sludge had a typical analysis as given in Table 12.

The overall response of the recycle process may be illustrated by comparing the product levels from the Densator with those resulting from the "Yellowboy" process with sludge separation by a thickener. A typical Proctor No. 2 water containing about 400 mg/l iron is the base for the comparison.

Table 11. DENSATOR RECYCLE TEST - OPERATING PARAMETERS

	Maximum	Minimum	Optimum
Raw Water			
pH	3.4	2.9	
Fe^{+2} - mg/l	530	100	
Total Acidity - mg/l	1,700	1,500	
Sulfate - mg/l	2,500	2,300	
Temperature - °C	18	8	
Densator Effluent			
pH	6.3	9.1	7.7
Fe - mg/l	91	<1	<1
Densator Feed			
Total Feed Flow Rate	160	50	40
Precipitable Solid (PS) - ppm	1,400	400	
Recycled Solids (RS) - ppm	20,000	60,000	30,000
Ratio RS:PS	14	150	30
Total % Solids in Primary Zone	2	6	3-4
Flocculant Addition - ppm	5	0	5-10
Recycle Slurry Flow Rate - gpm	60	10	10
Raw Water Feed Flow Rate - gpm	140	15	30
Ratio Fe^{+2}/Fe^{+3}	4:1		
Settling Rate - ft/hr[a]	2.3	0.3	1.0
Densator Underflow			
% Solids (average operation)	20	2	10-12

Densator Area required for optimum conditions -

ft^2/ton PPTD solids/24 hrs - 660

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[a] At solids concentrations of greator than 4%, the slurries are already at compression zone conditions; hence, free settling rates are not obtained.

Figure 13. Steady state densator flow rates for various recycle operating conditions

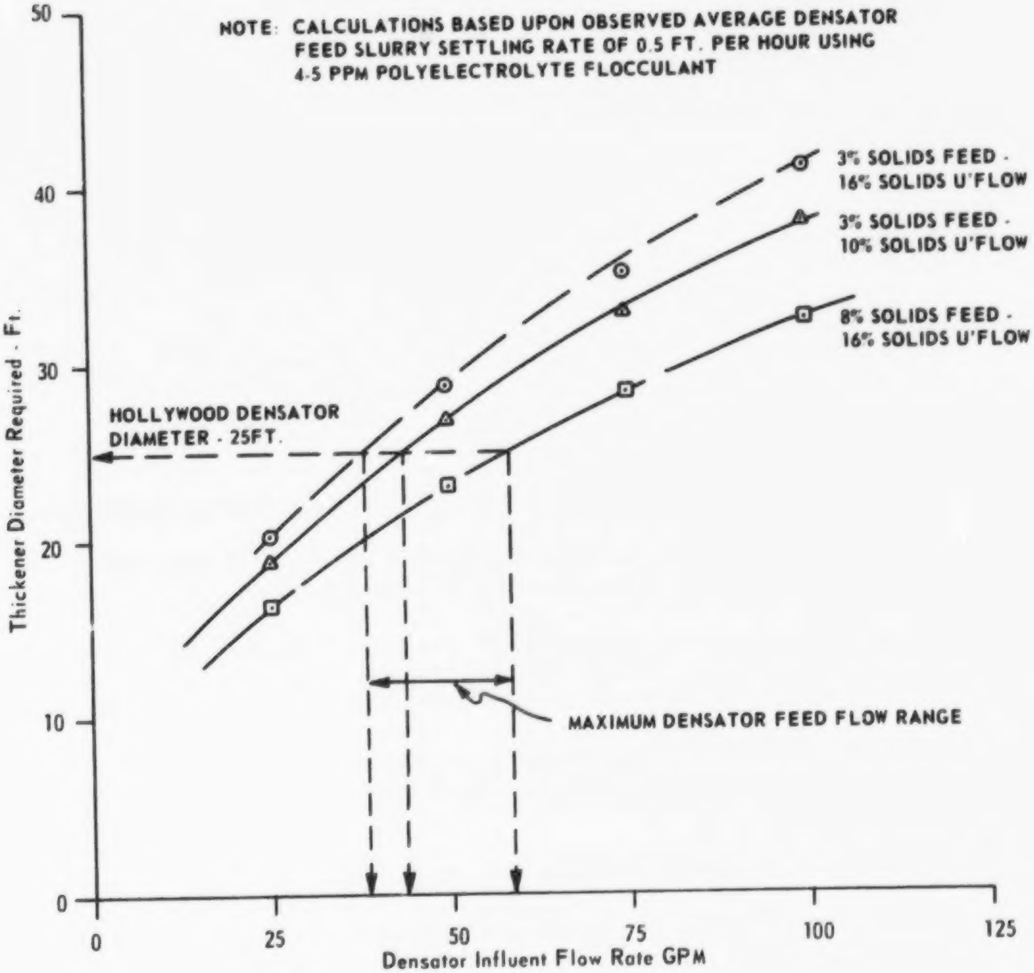


Table 12. DENSATOR RECYCLE SLUDGE ANALYSES TREATING PROCTOR NO. 2 DRAINAGE

		Calculated component composition	
Constituent	Weight %	Component	Weight percent
Aluminum	6.65	$CaSO_4 \cdot 2H_2O$	20.1
Iron	30.85	$Al(OH)_3$	21.2
Calcium	4.67	$Fe(OH)_3$(30% of Fe)	17.8
Sodium	0.02	$Fe(OH)_2$(70% of Fe)	35.0
Sulfate	12.06		
		Total	94.1

	Yellowboy thickener process[1]	Infilco densator recycle process[1], [2]
Plant Effluent gal/1,000 gal. feed	927	992.2
Sludge Production gal/1,000 gal. feed	73	7.8
Sludge volume reduction, %	--	89.0
Solids content of discharged sludge, %	2	16.0

[1] Assuming the neutralized feed to the thickener contains 1,500 ppm solids.

[2] Assuming recycle ratios of 30:1 by weight (recycle sludge/new sludge) and 1:4.25 by volume (recycle sludge/raw water).

Simple material balance calculations readily indicate the impracticality of the recycle concept when applied to weakly mineralized waters. The sludge recycle load to maintain 3-4 percent solids in the reaction zone would be unrealistic. The recycle system is susceptible to wide fluctuations of performance as feed water characteristics change due to rain or events in the originating mine.

CONCLUSIONS OF DENSATOR STUDIES

1. The Densator functions efficiently as a multipurpose unit to treat coal mine drainage and to develop a high density sludge.

2. The recycle treatment process functions most effectively when treating highly mineralized waters having a high Fe^{+2}/Fe^{+3} ratio. A decrease in this ratio and in total iron loading tends to decrease sludge density.

3. Sludge densities of 10 to 16 percent solids were obtained consistently when waters with high ferrous iron ratios are processed. Intermittent samples showed higher density levels.

4. The sludge produced has excellent dewatering characteristics as related to either vacuum filtration or sludge drying bed operation.

   The settling rate of the sludge found in the primary and secondary zone of the Densator is relatively low (about 0.5 ft/hr) and becomes the primary criterion for Densator sizing.

5. Under the conditions of this test, flocculants at high dosage levels were effective and necessary to maintain reasonable flow rates. Weakly anionic high molecular weight polymers were found most effective.

6. Densator area requirements are about 500-600 ft^2/ton precipitated solids per 24 hours.

7. Difficulties in maintaining high density underflows were experienced when wide fluctuations in total and ferrous iron content of the feed drainage occurred.

SECTION IX

TREATMENT OF COAL MINE DRAINAGE WITH CAUSTIC SODA

The tests using sodium hydroxide to treat drainage waters were carried out with the conventional "yellowboy" procedure (Process 10, see Table 15, Appendix C). Each of the four waters was treated at flow rates ranging from 87 to 292 gpm. As with the calcined lime studies, the water temperatures were low, frequently near freezing. The reagent was supplied as a 50 percent solution in 55 gallon drums. The undiluted reagent was added to the flash mixer and controlled manually through a variable-speed, positive displacement pump to a head tank with return overflow. The head tank provided improved flow control at very low flow volumes.

This reagent is dangerous to handle, requiring gloves and eye protection. The drums utilized to ship sodium hydroxide are not pressure tested. Safety considerations prohibit the use of air pressure to discharge the solution from the drums. This reagent freezes at relatively high temperatures (54°F), a factor which frequently required the drums to be heated to allow transfer of the reagent. Commonly, shipments were received in a solid condition. The reagent drums tend to develop a concentrated slurry in the bottom which can block the delivery system, create difficulties in total recovery of the reagent from the drum and lead to some deviation in concentration which complicates treatment control.

The reagent requirements were estimated from acidity and flow of the raw water with operating control adjustments made manually from pH measurements in the flash mixer and oxidation tanks. Automatic addition of caustic soda solution is quite feasible. During tests at large flow volumes (200 gpm) of highly mineralized waters, the consumption of reagent was so great that a full time helper was necessary to change drums. In operations utilizing more than one drum per shift, the labor costs to handle the reagent supply become excessive. In such instances large, expensive, storage facilities would be required. These requirements (capital costs and labor) for delivery, storage and handling of this reagent further add to its high utilization costs. However, the use of a purchased, aqueous reagent does minimize equipment and labor requirements for reagent preparation.

The detailed results of these studies are given in Appendix C, Tables 15 through 18. The reagent effectiveness and control capability was satisfactory while treating the lower mineralized waters (Tyler Run and Bennett's Branch). However, in the tests with the more severely polluted waters the tendencies observed with hydrated and calcined lime toward excess alkali consumption resulting from aluminum and magnesium reactions become pronounced. Utilizing manual control, the pH variation in the oxidation tanks with flow and acidity did not show the trend as did lime products under automatic control. The pH deviations were smaller (0.77-4.27 - compare with Table 10). Controlled, step-wise addition of caustic soda as described with lime product reagents is indicated. The extreme difficulty of maintaining a pH below 7.0, with acceptable iron and acidity levels in the effluent, was evidenced by these tests. Although the ease of control with sodium hydroxide is excellent, there is a danger of overtreatment with subsequent damage to receiving streams. This potential hazard requires constant personal monitoring or elaborate automatic monitoring and control equipment.

The reagent consumption tended to be less than theoretical (66 to 125%) with the low acidity waters while maintaining a pH in the low six levels. Under these conditions a small acidity value remained although final iron concentrations were satisfactory. The consumption excesses tended to increase with flow rates, contrary to the experience with lime products. With the higher acidity waters, the reagent consumption increased drastically (82 to 175%) in seeking to maintain an alkalinity value in the effluents. These large excesses beyond the measured alkalinity levels must represent side reactions involving aluminum and magensium. In test 19-71, the final effluent persisted with an acidity level (7 to 22 mg/l) despite a reagent consumption of 131 percent. The iron content of the effluent was consistently satisfactory. The effluents were sparkling clear, ranging between 4 to 22 mg/l suspended solids with turbidity levels between 2 and 10 jtu.

As with lime products, the sludge production increased with acidity of the feed waters. The density (volume/weight percent) of settled sludge (two hours) was somewhat lower than with lime products: Tyler Run - 0.15, Bennett's Branch - 0.25 to 0.30, Proctor No. 1 - 0.19 to 0.28, and Proctor No. 2 - 0.61 to 0.76. The dry solids production (lbs/1000 gallons) was definitely lower than with lime products: Tyler Run - 0.37 compared with 0.55, Bennett's Branch - 0.72 to 0.88 compared with 0.23 to 1.67, Proctor No. 1 - 2.16 to 3.28 compared with 2.7 to 4.9, and Proctor No. 2 - 10.1 to 12.8 compared with 11.7 to 19.1. The settled sludge volume (percent) was greater than with lime products for the low acidity waters (3.0 to 3.5) but less voluminous with the high acid waters (14.0 to 20.0), compared with a maximum of 29.6 with calcined lime. The sludges displayed settling behavior described as Type I or II depending on the raw water acidity. These sludges also tended to gel on standing. Table 24 provides sludge settling data. The shortest retention time utilized in the oxidation tanks, providing 0.28 minutes/ppm iron II, was adequate. The oxidation rates observed were similar to those with lime products. The oxidation-reduction potentials measured in the oxidation tank were somewhat higher than with lime reagents, especially in the lower acidity waters (128-540 millivolts).

The sulfate content of Tyler Run and Bennett's Branch effluents was reduced to less than 25 percent of the feed water. The conductivity of the effluents was doubled (171 to 258%) in the low acidity waters, with a smaller increase of Proctor No. 1 effluent (113%).

The treatment of CMD with caustic soda is characterized by its rapid reaction rate and quick responses, necessitating careful control. It is an expensive reagent, dangerous to handle, which may require special storage facilities. Its use eliminates reagent preparation equipment. The weight of sludge produced is somewhat less than formed with lime. This is attributed to the lack of precipitation of calcium sulfate. The sodium hydroxide added to the several waters as a neutralization reagent would result in a sodium concentration between 45 and 742 mg/1 (see Table 17, Appendix C). Tests 1-71 and 2-72 gave treated water effluent levels of 59 and 49 mg/1 respectively in contrast to sodium additions of 57 and 63 mg/1. Sodium levels found in the raw waters ranged between 4 and 49 mg/1 (see Table 2, Appendix A). Otherwise the overall response is similar to that observed with lime product reagents. Its greatest advantage lies with emergency use to prevent discharge of unacceptable quality waters.

SECTION X

TREATMENT OF COAL MINE DRAINAGE WITH SODA ASH

The reactions with sodium carbonate were carried out with its addition to the flash mixer as a dilute solution prepared in the dissolver (see Appendix B, Figures 8 and 9 and Table 20 in Appendix C). The high solubility of soda ash (7^{0° - 45^{100° gm/100 ml) permits concentration control. Although caution must be exerted, the handling danger is less than experienced with caustic soda. Unless fed as a solid, dissolution equipment is necessary. Each of the four waters was treated in flows ranging from 105 to 288 gpm. The water temperatures ranged from 38-45°F, excepting Proctor No. 2 (50°F), whose waters are collected in the mine. The system was controlled by a pH probe located in the first compartment of the flash mixer. Following iron oxidation, the solid-fluid separation was made in either the settling lagoon (Process 10) or in the thickener (Process 11). The data are presented in Appendix C, Tables 20 through 24 for Process No. 10, and in Tables 25 through 29 for Process No. 11.

The reaction control, as evidenced by pH measured in the oxidation tanks, was excellent with the extreme values occurring in less than 20 minute intervals. The improved uniformity results from the highly buffered system. In maintaining these pH levels, the response of the electrodes in the flash mixer led to the addition of excess reagent, developing conditions favoring unwanted reactions. The excess reagent did not remain available as free alkalinity. Even with the very large reagent excesses employed, the buffering effect did not allow the pH to increase to unacceptable levels. Improper dosages of soda ash can develop high pH levels, but these are less likely to occur than with caustic soda. The pH ranges in the flash mixer and oxidation tank for these tests were:

	pH Range	
Test No.	Flash Mixer	Oxidation Tank
25-71	6.0 - 8.5	7.0 - 8.2
26-71	7.5 - 9.6	7.3 - 8.1
29-71	5.0 - 10.0	6.8 - 7.0
27-71	7.8 - 10.1	7.3 - 7.8
30-71	6.4 - 10.6	7.0 - 9.0
28-71	8.2 - 9.6	7.8 - 8.3
31-71	5.8 - 9.8	6.6 - 8.2
32-71	6.3 - 9.8	7.6 - 8.4
34-71	7.6 - 9.2	7.2 - 7.6
33-71	6.6 - 9.0	7.3 - 8.3
36-71	6.9 - 8.6	7.5 - 7.7
37-71	3.6 - 9.6	6.9 - 8.0
35-71	6.0 - 9.5	6.3 - 8.0

The observed plant response was confirmed with a laboratory test using Proctor No. 2 water (1,420 mg/l acidity). Four liters of this water (requiring 6.0 grams of Na_2CO_3 based upon a neutralization equivalent of 8.8377 x 10^{-6} lb/gal/mg/l acidity) was neutralized with dry, CP laboratory reagent. Small increments of the solid reagent were introduced over a 30 minute period until the solution reached a pH of 8.3, simulating plant tests. The water temperature was 60°F. Vigorous mechanical stirring served to introduce air into the water for continuous oxidation. An addition of 10.7 grams sodium carbonate was required. This represented an alkali-use efficiency of about 56 percent.

In other laboratory tests, 400 ml of mine drainage (457 mg/l acidity) was treated by the slow, drop-wise addition of 0.192 gm (theoretical) of dry, CP, Na_2CO_3 dissolved in 20 ml of distilled water. The sample, at 65°F, was subject to vigorous stirring. After 30 minutes reaction time, the pH of the solution was 5.68. The addition of 50 percent more carbonate solution (96 mgm) increased the pH to 7.50, reflecting the buffering response. In another test with the same water, a 20 percent excess of Na_2CO_3 was reacted in the same fashion. The solution was then boiled (disrupting the bicarbonate-carbonate equilibria by driving off carbon dioxide) and cooled. The final solution had a pH of 7.18, 31 mg/l alkalinity and contained no detectable iron. Sodium carbonate will respond stoichiometrically with mine drainage when the carbonate equilibrium is controlled. However, such response cannot be expected in industrial practice. If used in mine drainage treatment, nearly 100 percent excess reagent consumption should be anticipated.

Soda ash in the form of solid pellets, sometimes called "prills", are used effectively in treating small water discharges during surface mining by immersing the prills, contained in a wire basket, in the draining stream. The slow rate of solubility from the prills tends to make treatment self-regulating. However, with high iron concentrations, the prills become coated with reaction products which prevent solubilization. This practice during surface mining provides a responsive, handleable reagent and avoids the mechanical difficulties associated with field use of powdered, hydrated lime. Low reagent efficiency results during such applications.

The reagent consumption experienced during the plant tests ranged from 146 to 313 percent. The excess increased with flow rates but showed no particular trend with acidity levels, which probably reflects the delayed response of the buffered system to pH changes. The effluents did maintain an alkalinity value but the percentage reagent utilized which was related to this contribution was small and large excesses of reagent went into the treatment per se, as shown below. The reagent representing effluent alkalinity decreases with increasing acidity of the feed water.

Test No.	% Reagent Feed Representing Effluent Alkalinity	% Reagent Consumed for Treatment Based on Theoretical
25-71	63.2	82.9
26-71	17.1	259.1
29-71	28.2	149.8
27-71	23.0	164.6
30-71	14.5	141.6
28-71	12.2	235.4
31-71	11.4	246.2
32-71	14.9	124.4
34-71	7.2	272.3
33-71	4.0	256.1

The iron removal response was excellent with the effluents having consistently less than 2 mg/l soluble iron. The effluents were clear with suspended solids ranging between 3 and 26 mg/l and turbidity levels between 2 and 20 jtu.

The sludge formed with soda ash settles to greater densities than sludges resulting from lime products but does not approach densities achieved with limestone. The two hour settled densities ranged from 0.37 to 2.1 weight/volume percent. The level of sludge production is equivalent to that found with caustic soda (0.25-19.4 lbs dry solids/1000 gal water treated) and less than the lime product sludges. The corresponding sludge volume was decidedly less than caustic soda and lime product sludges (0.3 to 11 volume percent) but was not as favorable as the response from limestone. The sludges may be described as exhibiting Type I behavior but have better mobility and are less inclined to gel. Typical settling rate curves are shown in Figures 24 and 25 with other sludge data shown in Table 24.

The retention time provided for iron oxidation was excessive, ranging from 0.78 to 72 minutes/mg/l iron II. The fastest oxidation rate was at least 230 mg iron II/l/hr. The oxidation-reduction potentials were favorable within the limits of 84 to 383 millivolts. The low iron levels in the effluents also reflect proper oxidation. The use of sodium reagent in excess resulted in substantial increases in dissolved solids as reflected by conductivity. The average increase of about 450 micromhos was substantial (from 42 to 254 percent). The sodium concentrations which would be developed in the water with the reagent excesses employed ranged from 55 to 1713 mg/l. This response was similar to that experienced with caustic soda treatment.

The treatment response of Proctor No. 2 waters when using the thickener (Process 11) for solid-fluid separation confirmed the other studies using the settling lagoon. The thickener settling operation was equivalent to similar tests with lime product treatment. The reagent consumption was excessive (223 to 292 percent). The effluents were alkaline with 1.4, 28.0, and 13.1 percent (Tests 36-71, 37-71, and 35-71, respectively) of reagent added occurring as alkalinity in the effluents. The actual excess reagent consumed in the reaction was uniform for the three tests: 219.8, 210.0, and 194.8 percent. The consistent effluent pH levels were between 7.56 and 7.71 reflecting good reagent control response. The pH variation during the three runs was 2.50, 2.86, and 0.90. The iron levels were satisfactory with an unexplained exception in test 36-71.

The two hour sludge densities were less than the Process 10 tests (0.56-0.75 weight/volume percent) as was the sludge production (5.1 to 6.9 lbs dry sludge/1000 gallons) and sludge volume (1.1 percent). This response is inconsistent since the raw water loadings were similar.

The oxidation retention times were less than the Process 10 studies (0.52, 0.49, and 0.39 minutes/mg/l iron II respectively for the three tests) but were adequate.

In all of the tests utilizing soda ash with the settling lagoon, the effluent specifications were achieved; however, the reagent consumption was excessive. Of the excess alkali, only a portion occurred as effluent alkalinity. The remainder apparently being consumed in other reactions, favored by the buffered conditions. The sludge characteristics are somewhat more favorable than those resulting from lime products or caustic soda, especially in resulting sludge volumes. The process drastically increases dissolved solids in the effluents.

SECTION XI

TREATMENT OF COAL MINE DRAINAGE WITH DOLOMITIC REAGENTS

There were three groups of studies made with dolomitic reagents considering the hydroxide, the oxide anhydride and the carbonate. Some process variations were also introduced. All four waters were treated at flow rates from 31 to 139 gpm. The data for the hydrated dolomite studies are given in Appendix C, Tables 29 through 33. The prepared hydroxide slurry was reacted in the flash mixer followed by air oxidation and solid-fluid separation in the thickener. The effluent was discharged through the settling lagoon. The pH control probe was located in the flash mixer with a set point of 7.0. The slurry preparation did not differ from the use of hydrated lime. The calcined dolomite values are given in Tables 34 through 39 of Appendix C. This reagent was utilized in the same flowsheet as the hydroxide, replacing the dissolver-mixer with a slaker. The calcined dolomite responded differently than the anhydride lime during slaking. The slaking temperature was difficult to maintain; the solid feed resisted wetting. A solid crust tended to develop in the slaker reaction chamber and formed a thick mix which moved with difficulty to the grit separator. Much dilution water was required to separate the grit. The hydration may have been inadequate, producing a less reactive reagent. This behavior was related to the inert, insoluble MgO component, since only the calcium portion of the compound slakes under atmospheric pressure. The dolomite was fed as larger-sized particles (Grade 4) to the rotary reactor. In some tests, it was supplemented with pulverized limestone. The data are presented in Appendix C, Tables 37 through 42.

The dolomitic hydrate yielded a satisfactory effluent with the highly mineralized Proctor No. 2 waters. The reagent consumption was reasonable although in one test a 146 percent consumption occurred. Even this level was less excessive than experienced with high calcium reagents.

The pH control of the reaction was good, maintaining a level above 7.5. The effluents had an excess alkalinity, which accounted for only 1.3 and 1.6 percent of the reagent fed in tests 58-71 and 60-71, respectively. When the reagent was introduced in large excesses, it did not appear as water alkalinity. The iron removal and effluent clarity were excellent. The settling lagoon effluents showed a reduced magnesium content, but the calcium levels rose nearly four-fold. It appears that magnesium was being removed from solution.

The settled sludge densities (0.58 to 1.3 weight/volume percent) were consistent with those resulting from hydrated lime treatment, as were the sludge production levels (11.4-16.2 lbs dry solids/1,000 gallons). The sludge volumes were somewhat greater (to 33 volume percent) but were subject to reduction under compression. Settling rate curves are shown in Figure 25, while other sludge data are given in Table 24. The settling behavior of these sludges respond as Type I, tending to gel. The sludge composition (Table 50) had unusually high calcium but low magnesium and sulfate content. Sludge drainage rates were also consistent with other lime product reagents as shown in Table 61. As discussed in Section XII there is strong evidence for magnesium-sulfate-complex association, indicating possible advantages for magnesium reagent in treating high sulfate-containing waters. Response of these sludges in the thickener and to filtration is cited subsequently in reference to the calcined dolomite tests.

The retention time utilized during iron oxidation (0.61-0.90 minutes/mg iron II/l) was probably excessive. The oxidation rates exceeded 263-297 mg/l/hr, while the oxidation-reduction potential ranged between 202 and 229 millivolts. The settling lagoon maintained a pH consistent with the oxidation tank effluent.

The calcined dolomite also gave satisfactory responses with all four waters using low flow rates. It appears that stoichiometric dosages of calcined dolomite are feasible, assuming reaction with only the calcium portion of the compound. The reagent consumption levels were consistent, with near theoretical values being attained for tests 61-71 (low acidity water) and 64-71 (high acidity water at a low flow rate). However, tests 62-71 utilized 186 percent theoretical reagent while test 63-71 was atypical (605 percent excess), being unexplainably out-of-control. These last two responses must relate to reaction time in the flash mixer, pH probe location or possibly to unobserved mechanical problems.

In test 63-71, utilizing the control set point of 7.0, reaction control difficulty was experienced throughout the test with the pH in the oxidation tanks rising to 9.5 for over three hours. The system was mechanically and chemically responsive. The control point was lowered in increments to 5.3 with the oxidation tank effluent pH slowly declining to 8.0 and finally to 7.5 near completion of the test. The reagent feed rate also approached theoretical with the control change. The test illustrated control problems which could be costly in unattended plants and lead to dangerous overtreatment with reactive reagents.

These tests showed greater variation in the alkalinity of the oxidation tank effluent than with the hydrate reagent. Each test having reagent excesses did not develop an alkalinity. The two (61-71 and 64-71) attaining an effluent alkalinity represented 15.6 and 4.2

percent of the reagent fed. The iron reduction was satisfactory. Reactions with these reagents continued throughout the process as there was a rise of pH levels in the settling lagoon effluents.

The settled sludge density confirmed the values from the hydrated dolomite tests, as did the sludge production level. The settled sludge volumes were less than resulted with hydrated dolomite and more consistent with lime product reagents. The sludges showed Type I settling behavior and would gel. Other sludge settling data including thickener observations are given in Table 34. The chemical analyses of these sludges used in filtration tests are given in Table 49. They are high in calcium and magnesium but generally lower in sulfate content. The filtration data generally agrees with that from lime product sludge (see Tables 51 and 55). Sludge drainage rates (drainometer) are reported in Table 60 indicating their bulk densities to be slightly lower than hydrated lime sludges.

The iron oxidation levels were satisfactory at the retention times utilized (0.76-82 minutes/mg iron II/1), covering the range 3 to 303 mg iron/1/hr. The oxidation-reduction potential ranged between 197 and 293 millivolts.

The dolomite tests indicated this reagent to be less responsive than high calcium limestone, but capable of developing higher pH levels in the effluent. With the time allotted to study this reagent, it was not possible to explore all the parameters of the rotary reactor and stone properties (only one particle size was used) which control the reaction. The reagent provided adequate treatment for Tyler Run waters in the rotary reactor at a flow of 89 gpm. The feed water had little iron (< 2 mg/1) which made iron oxidation unnecessary. The reactor effluent maintained a pH above 7.0, a reduced iron content, and an excess alkalinity. The alkalinity added to the effluent represented 58 percent of the neutralizing requirement. The estimated reagent consumption was excessive since equilibrium conditions were not established. This unreacted reagent was not solubilized but reported to the sludge. With appropriate control, reasonable consumption could be expected. Under these conditions reagent cost would not exceed 2 cents/1000 gallons.

The sludge behavior responded as those produced in limestone reactions (Type III) maintaining a slight turbidity. The sludge settled rapidly to a high density (9.1 weight/volume percent) and small volume (0.4 percent). The sludge production was 3.0 lbs dry solids per 1000 gallons of water. This sludge contained excess dolomite fines. This sludge formed from a reactor effluent containing 600 mg/1 suspended solids. The settling lagoon effluent did not change in quality from the reactor effluent, indicating little continuing reaction. This response was different than observed with high calcium limestone.

The reduced reactivity of dolomite was indicated in test 51-71 when the flow rate was increased about 50 percent. Despite the available reactive reagent surface experienced in test 50-71, effluent quality was not maintained (pH 5.7) although 85 percent of the acidity was neutralized. The suspended solids in the reactor effluent also decreased. Subsequent experience has indicated satisfactory treatment is feasible by larger reactor loadings and control of reagent particle size. In this test, an acceptable effluent was attained by the introduction into the reactor of a dilute, pulverized, high-calcium, limestone slurry--thus utilizing a mixed reagent. The effluent pH increased to above seven and an alkalinity developed. The introduced pulverized limestone accounted for 78 percent of the reagent requirements, the remainder being developed from the dolomite, showing further evidence of the relative reactivity between these two reagents. The reactions continued in the settling lagoon, as previously experienced with limestone products. The resulting sludge production was reduced (1.1 lbs dry solids/1000 gallons), as was settled sludge density (3.4 weight/volume percent) but the sludge volume remained constant. Flocculant addition was successful in improving effluent clarity.

Confirmation of these results using mixed reagents was attained in tests 53-71 and 54-71 with Bennett's Branch drainage of slightly different composition than Tyler Run. However, when a higher acidity and iron II water (Proctor No. 1) was treated, the reactor effluent pH was below seven and maintaining alkalinity was inconsistent. Only about half the ferrous iron had oxidized and been removed from solution. Additional oxidation time was necessary which was accomplished by transferring the reactor effluent to the oxidation tanks to provide a long retention time (10.2 and 4.2 minutes/mg iron II/l, respectively for the two tests). This increased response time reduced the iron levels satisfactorily. The mild agitation in the oxidation tanks also favored continuing neutralization reaction which allowed development of an alkalinity in the treated waters. The level of agitation in the oxidation tanks allowed some sludge deposition with a resulting decrease in suspended solids, as in test 54 from 1300 to 571 ppm.

These studies show the reduced neutralization reactivity level of dolomite in respect to high calcium limestone. Waters with acidity levels above 300 mg/l may be treated satisfactorily but only with great increases in reaction time. As cited in Section XII the reaction times necessary with relatively insoluble carbonate reagents is extensively controlled by those parameters with establish contact between available reactive solid surface and diffusion levels control carbon dioxide concentrations. The conditions maintained for these tests utilizing the rotary reactor were inadequate when utilizing dolomite but were satisfactory when applied with high calcium limestone. The use of supplemental reagents, as pulverized limestone, can be helpful in improving these reaction times.

The evaluation of dolomite-based reagents indicate their potential use in treating coal mine drainage but point to major limitations. In Pennsylvania, the higher cost of these reagents (mainly transportation costs) would discourage their use; however, in locations where the reagent supply sources are closer to the treatment plant, their use may be indicated. Under good control, a near stoichiometric consumption of the hydrated and calcined products can be expected based upon reaction of the calcium component. The more costly pressure-hydrated product could be expected to achieve full utilization but was not tested. The treatment effluent will meet requirements but overall reaction control appeared more difficult to attain. The effluents tended to maintain a slight turbidity which can be controlled with flocculants. The natural dolomitic carbonate was less reactive than high calcium limestone but did permit attainment of higher pH levels. As with limestone, any unreacted reagent reported to the sludge having minimum water solubility. The use of mixed reagents offers advantages. Further study should evaluate combined calcium-magnesium reagents to minimize the gypsum deposition problems in high sulfate-containing drainage.

SECTION XII

TREATMENT OF COAL MINE DRAINAGE WITH LIMESTONE

The tendency to oversimplify processes to treat coal mine drainage has led to overemphasis of the ratio of reagent cost to overall treatment economics. Yet the selection of the reagent establishes the system and to a significant extent controls total treatment costs. The theoretical 1.92 cents per neutralization equivalent for limestone, with its 2.56-fold advantage over quicklime and 3.83-fold advantage over hydrated lime (see Table 2) has continued to be viewed advantageously. Despite this and other advantages, pragmatic process considerations have prevented the more extensive use of limestone. These studies sought to develop a perspective of this situation.

Attempts have been made to use limestone. Tracy[16, 17] in 1916 utilized -200 mesh pulverized stone seeking an economic product from the sludge. Zurbuck[18] in 1963 developed a water-powered drum reactor, filled with limestone, to maintain alkaline quality of a remote forest stream. Researchers at the U.S. Bureau of Mines[8, 19, 20] subsequently considered an electric powered, three-foot diameter, 24-foot long rotary reactor using limestone for short duration tests. The necessity for additional iron II oxidation and carbon dioxide removal was reported. They favored using the reactor as an autogenous mill. Most significantly, they reported the settled sludge volume to be one-third that resulting from hydrated lime. Stone-use efficiency of about fifty percent was suggested. The rotary reactor is not an efficient comminution device; the grinding response can be improved by a ball mill operation[21]. The greater reaction efficiency from using -400 mesh, high carbonate stone as well as response to other stone properties have been reported[22, 23]. Such fine-sized stone is not prepared commercially. The available pulverized product (-200 mesh) requires long reaction times. At some quarry locations, large stocks of ungraded fines are available as waste by-products. To attain desired reagent efficiency during CMD treatment, the recovery and recycle of unused reagent is necessary. The low cost of limestone has not suggested such recovery to be realistic.

In 1968 Calhoun[24, 25] described satisfactory results from the only mine plant location using stone in a rotating reactor which was fed with ferric waters. Stone hardness was reported to be the most important parameter. Birch[26, 27] reported combined use of stone and hydrated lime. Holland, et al.[28] reported pilot plant studies using a tumbling mill to produce limestone fines for neutralization which was followed with aeration. Hydrated lime was necessary to complete the treatment. These authors concluded aeration costs following limestone neutralization are impractically high.

More recent pilot studies by Wilmoth, et al.[29, 39] utilized -200 mesh pulverized limestone to treat ferric waters, and further confirmed the continuing slow reactions. Excessive stone quantities were required. Other valuable studies with limestone were reported by Gehm[30] using an upflow stone reactor system to treat pickle liquor; Hoak, et al.[31] also treated pickle liquor with limestone. Other authors studying limestone neutralization reactions include: Khavskii et al.[32], Jarrett and Lountz[34], Yen[35], Clifford and Snarley[37], Armco Steel Corporation[38], and Glover, et al.[40, 41, 42].

CHEMICAL ASPECTS OF LIMESTONE TREATMENT OF COAL MINE DRAINAGE

In the general equations written for limestone neutralization of coal mine drainage, the water solubility relationships existing for the reactant and products are too often ignored:

$$CaCO_3 + H_2SO_4 \rightarrow CaSO_4 + H_2CO_3 \qquad (7)$$

$$3CaCO_3 + Fe_2(SO_4)_3 + 6H_2O \rightarrow 3CaSO_4 + 2Fe(OH)_3 + 3H_2CO_3 \qquad (8)$$

$$3CaCO_3 + Al_2(SO_4)_3 + 6H_2O \rightarrow 3CaSO_4 + 2Al(OH)_3 + 3H_2CO_3 \qquad (9)$$

The small amount of $CaCO_3$ (0.014 gm/l), which dissolves in water, reacts rapidly with mine water. The continued neutralization reactions are governed by the limestone solubility rate which varies with the physical properties of the stone, the presence of dissolved salts, and the carbon dioxide partial pressure. There is also some reaction at the solid limestone surface which is governed by surface area, surface reactivity, diffusion rates of reactants and products, and reactant concentrations. Ohyama, et al.[43] considered this mechanism. As the pH rises above 4.5, the equilibrium which yields an excess of hydroxyl ions become increasingly important. In this same pH region, the solubility products of the hydrated ferric and aluminum compounds are exceeded and their precipitation begins. Ferrous compounds are not precipitated until higher pH levels develop. Calcium sulfate, another reaction product, also has limiting solubility and plays a unique role in this process. Gypsum tends to deposit and grow, though its solubility may not be exceeded. Localized concentrations of gypsum, especially on the limestone surface, are undoubtedly high. Carbonic acid, subject to dissociation, acts as a buffer to maintain a weakly acid environment. Carbon dioxide can be stripped from such waters allowing the pH to rise.

These equilibria suggest overall slow, continuing responses, especially when fine (near-colloidal) calcium carbonate particles are present. They also indicate why mechanical effects tend to become controlling.

The overall chemistry of this system is complex, especially the solution chemistry of iron between pH 2 and 10. Of concern are: the solubility products of iron, calcium and magnesium salts (sulfates, carbonates, bicarbonates and hydroxides) as well as carbon dioxide and oxygen; the interactions of three physical states (and the response of partial pressures, complexes, ionic strength, particle size, nucleation); and reaction conditions (temperature, contact times, concentrations, oxidation state, equipment configurations).

Some appropriate solubility data are cited which indicate a wide range for the Ca^{+2} and SO_4^{-2} ions as a function of carbon dioxide partial pressure and the solid phases: calcite, dolomite and gypsum. Shterning and Frolova, 1945[12], give solubilities of the $CaCO_3$-$CaSO_4$-CO_2-H_2O system, while Frear and Johnson, 1925[12], give more detailed data at higher partial pressures of carbon dioxide. Cameron and Seidell, 1906[12], indicate a negligible effect of calcium hydroxide on the solubility of gypsum. Gypsum solubility is not effected by low concentrations (1000 ppm) of $Ca(HCO_3)_2$, but may be decreased at higher levels. Table 13 describes the CO_2-$CaCO_3$ system. Calcite solubility is largely a function of concentrations of HCO_3- ion, and the partial pressure of carbon dioxide. Table 14 indicates the variation in gypsum solubility with temperature. Table 15 shows the response of magnesium on calcium carbonate solubility. Yanateva's (Table 15) and Garrel's[45b] data both suggest a complexing effect between magnesium and sulfate ions which enhances solubility of the latter. This effect suggests differences between the use of dolomite and calcium carbonate. In particular, $CaSO_4$ precipitation may be significantly reduced when using dolomite. This has significance when treating severely loaded CMD. These waters often cause difficulty by the precipitation of large amounts of gypsum which scales pipes, reaction vessels, agitation blades, etc. Gypsum coating should not be anticipated with waters containing less than 1400 ppm SO_4^{-2}. Ferric salt precipitation is independent of calcium salt phenomena.

To produce a pH stable plant effluent in CMD treatment, the neutralization of free acid and hydrolyzable salts with their subsequent precipitation must be complete and the excess carbon dioxide eliminated. This stability must include the sulfate and bisulfate complexes and acid generated by the iron and aluminum salt hydrolysis. Ionic forms of some iron compounds typical of CMD, and their equilibria formulae, are cited by Pourbaix[44]. An adapted "potential-pH" equilibrium diagram for the limestone-iron neutralization system is shown in Figure 14. By proper control of pH and Eh, it is possible to form ferrous hydroxide, ferric hydroxide, ferrous carbonate or mixtures thereof. The region designated "A" in Figure 14 represents typical treatment conditions, but the condition could be extended to include region "B" while using hydroxide reagents. With limestone systems a

Table 13. SOLUBILITY OF CALCIUM CARBONATE IN WATER IN CONTACT WITH AIR (16°C)
(Combined data of Schlossing, Engel, and Johnston-**Seidell** and Linke)[12]

Partial pressure of CO_2 in atmospheres	Total calcium ppm	Total $(HCO_3)_2$ ppm	Total $Ca(HCO_3)_2$ ppm	Total $CaCO_3$ Equivalent ppm
0.0005	30	89	121	75
0.0008	34	136	160	85
0.0033	55	166	221	137
0.0139	90	271	361	223
0.0282	119	361	480	296
0.0500	144	439	583	360
0.1422	214	648	862	533
0.2538	265	808	1,073	663
0.4167	313	960	1,273	782
0.5533	354	1,080	1,434	885
0.7297	391	1,190	1,581	997
1.0	432	1,325	1,957	1,086
2.0	566	1,725	2,291	1,411
4.0	734	2,240	2,974	1,834
6.0	855	2,600	3,455	2,139

Range of Ionic Strength (0.0005 to 1.0 Atm CO_2) = 0.0022 to 0.0324.

Increase in Calcium Solubility (0.0005 to 1.0 Atm CO_2) = 432/30 + 14.4 X.

The usual partial pressure of carbon dioxide in the atmosphere is 0.00034.

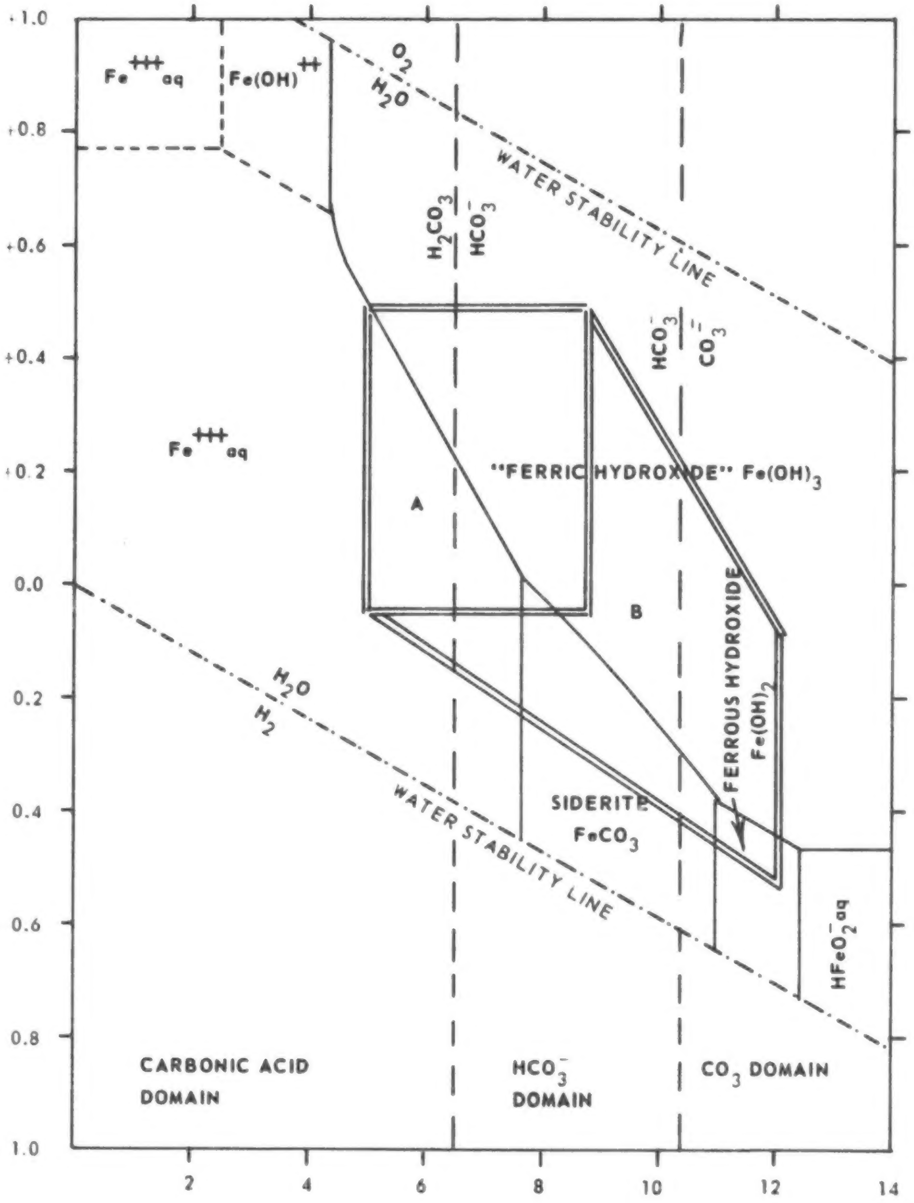
Table 14. SOLUBILITY OF CALCIUM SULFATE DI-HYDRATE ($CaSO_4 \cdot H_2O$ - GYPSUM) IN WATER (Hulett and Allen, 1902)[12]

Temperature °C	$CaSO_4$ ppm	Ca equivalent ppm	SO_4 equivalent ppm
0	1,759	517	1,242
10	1,928	568	1,360
18	2,016	595	1,421
25	2,080	612	1,468
30	2,090	615	1,475
35	2,096	617	1,479
55	2,009	592	1,417
100	1,619	477	1,142

Table 15. SOLUBILITIES IN THE QUATERNARY SYSTEM $CaCO_3 + MgSO_4 \rightarrow CaSO_4 + MgCO_3$ (Yanat'eva, 1955)[13]

Solubility of ions in ppm (at 0°C and 1.0 atm partial pressure of CO_2)				Solid phases present
Ca	Mg	$(HCO_3)_2$	SO_4	
600	0	1,860	---	Calcite ($CaCO_3$)
214	130	1,330	---	Dolomite ($CaCO_3 \cdot MgCO_3$)
545	32	1,850	---	Calcite and Dolomite
---	540	2,790	---	$MgSO_4 \cdot 7H_2O$(M)
136	295	1,940	---	Dolomite + M
970	---	1,660	1,040	Calcite + Gypsum ($CaSO_4 \cdot 2H_2O$)
1,040	39	1,900	1,170	Dolomite + Calcite + Gypsum
825	172	1,810	1,370	Dolomite + Gypsum
760	252	1,890	1,360	Dolomite + Gypsum
800	276	1,940	1,530	Dolomite + Gypsum
790	320	2,000	1,560	Dolomite + Gypsum + M
610	---	---	1,480	Gypsum

Figure 14. A limestone-iron neutralization system



pH of 8 is seldom exceeded. However, the location of the siderite field has potential of precipitating a more dense form of iron without aeration. The area of the siderite field can be increased with dissolved carbonate concentration and partial pressure of carbon dioxide.

Freshly precipitated siderite oxidizes more rapidly than ferrous ion in solution. If siderite were precipitated in the absence of air in a slightly reducing system, then aerated, a more desirable reaction sequence could develop. The rapid oxidation of insoluble ferrous compounds agrees with the observations of Lovell and Stauffer[46].

Similar equilibria data for the sulfur-water system[44a] and iron-water system (non-carbonate)[44b] are available.

Considerations of the characteristics of the impurity sludges are presented in Section XIV, including: solid-fluid separation, secondary dewatering, and disposal. The sludge characteristics are determined by the reagent and processes used in the treatment. Limestone treatment appears to result in an optimum situation forming a sludge which is rapidly settled, dense, and readily dewatered. The treated water tends to maintain a persistent turbidity after solid-fluid separation which clears very slowly. This haze is attributed to the presence of near-colloidal limestone particles, with the particles eventually settling or solubilizing. It is also feasible that part of the turbidity originates from colloidal silica in the limestone. The mass of sludge produced during treatment relates to the pollution loading of the raw waters, primarily the iron and aluminum content. In some high sulfate (> 1400 ppm) waters, gypsum may also form. Other sludge contents result from insoluble impurities in the reagent and from excess reagent. The inefficient use of limestone increases the mass of sludge to be handled and total process costs.

There is no reason to anticipate limestone requirements in excess of the stoichiometric value: 8.3453×10^{-6} lbs of calcium carbonate per mg/liter acidity per gallon of water treated. This value must be adjusted to consider the stone purity and the efficiency with which it is utilized. The later is dependent upon a myriad of parameters to be indicated from these studies.

CONSIDERATIONS IN THE USE OF LIMESTONE

The chemical factors of concern include: reactant solubility rates, precipitation and neutralization rates, iron oxidation rates (either pre- or post-neutralization), and composition of the sludge, which determines suspended solids settling rates, and sludge dewatering rates. These factors determine equiment selection and sizing, hence capital costs. The equipment basically establishes reagent efficiency and operating costs.

Among the advantages in the use of limestone as a reagent are:

1. Lowest cost neutralization reagent - about 0.002¢ per part acidity per thousand gallons.
2. Minimal sludge volume is developed - usually less than one-third volume from more soluble reagents.
3. The limestone-produced sludges have high solids content.
4. The sludge tends to be granular and crystalline, making dewatering operations less costly and more efficient.
5. It is impossible to overtreat the water, leading to minimum control problems.
6. Reagent availability is extensive over a wide geographical distribution.
7. Limestone storage is favorable. The larger sizes do not require protective enclosures. Pulverized stone, which requires bin storage, handles well and is non-hygroscopic.
8. Simplicity of plant design.
9. Reduced plant maintenance and operation requirements.
10. Reduced safety hazards relative to reactive reagents.
11. Reaction capability indicate stoichiometric reagent requirements need not exceed 20 percent theoretical.

There are limitations and disadvantage in the use of limestone. Among them are:

1. Use is limited by the ferrous iron concentration in the feed water. The limiting pH attainable, near 8.3, establishes maximum iron oxidation rate.
2. Reactivity is lower than more soluble reagents. Surface coating of the reagent by products reduces reactivity.
3. Settled sludge has clay-like character which may create handling difficulties.

4. The actual sludge weight produced (on a dry basis) may be greater than with lime products since stone tends to be more impure. Slower reactivity and lower use efficiency may introduce unreacted reagent to the sludge.

5. Near-colloidal-sized particles of stone and silica create turbidity in plant effluents.

6. The limestone-produced sludge decreases in volume very little under compression.

7. The carbon dioxide formed during reaction must be removed to provide a stable pH.

8. Rotating reactors are noisy. They require power to turn the drum and are inefficient in comminution.

The response of mine drainage waters to limestone treatment suggests three types of feed waters, grouped by ferrous iron content. They are listed in Table 16. The ferric iron acidity concentrations of CMD do not appear to be limiting criteria. In other tests, not detailed here, it was established that mine drainage containing up to about 100 mg FeII/l could be satisfactorily treated in the rotary limestone reactor at maximum flow without a separate iron oxidation stage. The stoichiometric oxygen requirements for the oxidation of this level of iron concentrations are closely met by the air saturation level in water, assuming no replacement by additional oxygen to the solution. However, at higher FeII levels, acceptable iron concentrations in the settling lagoon are not attained.

Various limestone reactor modes have been considered by Glover[42]. Two modes of reacting limestone with CMD have been utilized which differ primarily in feed stone size--pulverized or larger sizes (+ 1/4 inch). The pulverized reagent is mixed with the CMD in a flash mixer in near stoichiometric quantities. Long reaction times are required with vigorous agitation to maintain particle suspension. The equipment should be housed. This mode tends to result in poor reagent utilization.

Larger particle-sized reagents are introduced, in excess, to autogenous, attrition reactors. Only rotary reactor designs seem feasible. The reaction time is short. The reactors provide vigorous movement of stone and water but do not require weather protection or housing. Much of the water movement results from gravity flow. Reactor capabilities must permit desired reaction levels but avoid excessive loss of unreacted stone.

Table 16. POTENTIALITIES FOR LIMESTONE TREATMENT OF MINE DRAINAGE WATERS

Ferrous concentration mg/l	Response to limestone treatment
0-100	Effective treatment may be achieved without pre- or post-neutralization iron oxidation.
100-500	May be effectively treated but requires post-neutralization aeration and significant reaction-retention time. Pre-neutralization oxidation can reduce iron II concentrations.
>500	Potential treatment is uncertain with experience to date, unless combined with pre-neutralization ferrous iron oxidation to achieve the above ferrous levels.

Reagent cost favors the use of larger-sized stone. Although the reactivity increases with decreasing particle size, the extent each particle responds lessens. This phenomenon is associated with particle surface coatings by reaction products and, the ease in maintaining the surface active. With the commercially-available pulverized stone sizes, reaction times in excess of twenty minutes become essential to minimize loss of unreacted stone.

The power requirements between the two modes are utilized differently. With the pulverized stone, the power required to maintain particle suspension must also move large volumes of water which serves minimal useful functions. With the larger stone reactor systems, the power is more effectively utilized to create and maintain reactive surfaces.

The rotary limestone reactors perform two principal functions, the relative extent of each has not been defined: a) maintain freshly exposed, reactive surfaces by attrition to remove insoluble reaction products; and b) achieve an autogenous comminution of the stone to produce fines of great reactivity. Additional functions also occur: c) contacting the neutralization reactants to enhance reaction rates at the solid-liquid interfaces which includes wetting of solid surfaces; d) contacting the oxidation reactants (dissolved iron II and oxygen); e) contacting air (oxygen) and water to maintain an oxidizing environment; f) removing carbon dioxide from the water; g) controlling the reaction-retention time; h) transporting the water through the reactor, i) providing a suitable transport and distribution of the stone in the reactor to optimize the other functions.

The capability with limestone treatment to achieve an effluent with an alkalinity value is the main control requirement. The occurrences which do not achieve an alkaline effluent or tend toward inadequate reagent utilization may be modulated, but tend to have counteracting responses. They may be enumerated:

A. In pulverized stone systems:
   1. Increase reagent addition.
   2. Decrease stone particle size.
   3. Use stone of greater purity and reactivity.
   4. Increase reaction time.

B. In larger-sized stone systems:
   1. Increase rotation speed.
   2. Increase stone loading of reactor.
   3. Add attrition devices to reactor.
   4. Use smaller top size stone.
   5. Use softer stone.
   6. Use stone with higher alkali equivalent.

Should these steps fail to be adequate, supplemental additions of a more reactive reagent may be employed after the limestone reaction. A greater reagent efficiency may result by achieving only partial neutralization with limestone and completing the treatment with hydrated lime slurry.

Even with an adequate effluent alkalinity, a depressed pH may result from carbon dioxide solution. This gas should be stripped from the effluent by subsequent mixing or aeration. During this gas stripping, the alkalinity and pH of the waters may decrease but an excess alkalinity level must be maintained. Should an acidity develop, inadequate available alkalinity has been introduced to the water.

Should an excess ferrous iron concentration persist, the following actions may be considered:

1. Increase oxygen transfer rate. This dissolved oxygen level in the effluent should be maintained at least 5 ppm.

2. Increase reaction-retention time. Iron II oxidation rates are relatively slow at the pH levels prevailing under these conditions.

3. Increase pH of the system by the addition of further limestone or other alkali supplement.

4. Add a chemical oxidant such as chlorine gas or calcium hypochlorite.

5. Decrease ferrous iron concentration of feed water by dilution or pre-neutralization oxidation.

With the attainment of suitable water quality after separation of the sludge, effluent turbidity may persist. This can be controlled with polymeric flocculants if necessary, but it is usually non-deleterious.

PLANT OPERATING OBSERVATIONS

Limestone neutralization studies were made with the rotary reactor employing feed stone with varied top sizes from 1/4-inch to three-inches (see Appendix B, Figure 16). In some tests pulverized limestone was also added to the reactor as a slurry. In other studies pulverized stone slurry was reacted in the flash mixer and the Densator. Tests were made with high calcium as well as dolomitic stone from several geographical origins and were applied to the several waters available.

The rotary reactor was designed to meet the multiple functions cited (see Appendix B, Figures 6, 7, and 16). Vigorous movement of water and stone was achieved within the reactor by horizontal lifters which enhance autogenous comminution, mixing, and gas exchange. The rotational movement of the drum carried the stone and water load to about "ten o'clock" before discharge.

The 2 mm wedge wire screen, covering the discharge aperture, was effective in preventing loss of all but the finest size stone particles. The most excessive loss of unreacted stone, as well as potential sanding problems in pipes, was prevented by this screen. Some of the fine particles which do escape were trapped in the effluent sump while others are carried with the effluent flow, providing additional reaction time while reaching the settling lagoon. Physical characteristics of limestone and dolomite sludges found in the bottom of the settling lagoon, as the water level was dropped, are given in Table 17.

The sample stations may be located in Figure 11, Appendix B. The larger particles would be expected to settle in the first region of the lagoon, with decreasing diameters as distance from the weir increased. The particle sizes were small and their distribution relatively uniform. The bulk density and solids content were surprisingly high - showing a significant density increase under compression in comparison with 24-hour settled sludge (Table 24). X-ray diffraction patterns of these sludges showed calcite, dolomite, quartz, and

Table 17. CHARACTERISTICS OF LIMESTONE SLUDGE FROM SETTLING LAGOON

		Sample 1	Sample 2	Sample 3
Sample location Inches from weir		96	104	124
Bulk density - gm/ml		1.167	1.139	1.171
Solids Content Wt. % @ 105°C		35.4	30.2	37.9

Size Content Direct Weight Percent

Size (mesh)				
passed	retained	Sample 1	Sample 2	Sample 3
	100	2.2	1.4	4.7
100	200	23.2	39.1	26.3
200	325	24.5	47.9	25.2
325	400	19.2	8.4	27.4
400		30.9	3.2	16.4

and opal (α cristobalite), in addition to ferric oxyhydroxides. No difficulty was experienced from the build-up of non-reactive inerts in the reactor.

The systems used with pulverized stone permit no control over loss of unreacted reagent; the stone reacts, settles in the reactor or is carried with the water to the next unit operation. In two years of intermittant operation of the rotary reactor, there was no evidence of internal wear and corrosion. Rubber, ceramic, or other abrasion-resistant liner material would further extend reactor life. The control anticipated by varying the rotary reactor slope was ineffective due to inadequate design of the feed end of the reactor. The "open" design, which permits the extension of a vibrating feeder and the influent pipe into the reactor, also permitted excessive splashing from the reactor at low slope angles and high flows. The result was potential corrosion, and limitations in the useful limit of fluid flow and horizontal reactor angle. The horizontal reactor angle effects stone movement within the reactor, but has negligible control over water retention time. The angles available favored excessive movement of the stone to the exit end of the reactor. The reactive capacity of the device is optimized only if the stone charge maintains a uniform distribution throughout the length of the unit. The experience attained with this unit suggests a horizontal, or even a shallow reverse angle to be preferrable. The slope variation capability provided in this research reactor is not needed in an industrial unit. A minor design change to a "closed" feed system would circumvent these difficulties and enhance reactor performance.

The retention time within the rotary reactor is indicated in Figure 15 as measured by batch introduction of a soluble salt and monitoring of the effluent conductivity as a function of time. The median retention time appears at about 150 seconds regardless of rotation speed, flow rate (up to 250 gpm), or reactor angle (3-4°). The retention time in the pulverized limestone reactors is a function of the reactor size and flow rate.

The belt drive of the reactor created difficulties in "training" the belt. Changes in temperature, humidity, precipitation and load caused the belt to creep, resulting in unreliability and undue operator attention, thus limiting the testing program. A chain drive is preferrable.

The power requirements for the rotary reactor are distributed between those required to turn the base reactor, to move the water, and to abrade the stone. These are detailed in Figure 16. The power requirements associated with the water movement do not appear to vary appreciably with flow. They increase with stone loading and change rapidly with reactor rotation speed between 9.6 and 16 rpm. With this unit, the power costs range between 6 and 14 cents/hour or between 0.4 and 2.3 cents/1,000 gallons of water for the conditions cited. Power consumption by the reactor reflects current (amperes) levels drawn. These levels were monitored to protect the motor from overload and also served as a rough indicator of reactor stone loading. The range observed was from 12 to 24 amperes. More effective power transmissions would be expected with a chain drive.

In contrast, the two small mixers used in the flash mixer unit with pulverized stone were each powered by a 1/2 HP motor and would require about 1.1 KW/hr or between 0.18 and 0.32 cents/1,000 gallons of water. However, these mixers were designed to produce homogenous solutions from two liquids (mine water and alkali slurries) and did not provide adequate agitation to maintain pulverized limestone in suspension. The flash mixer had a median retention time between 2 and 15 minutes which was inadequate for pulverized limestone reactions. It is probable that the power requirements of a properly selected agitator in a flash mixer providing adequate retention time for use with pulverized stone would be comparable or exceed those of the rotary reactor since large volumes of water would be continuously moved at high velocities.

Figure 15. Limestone reactor retention time

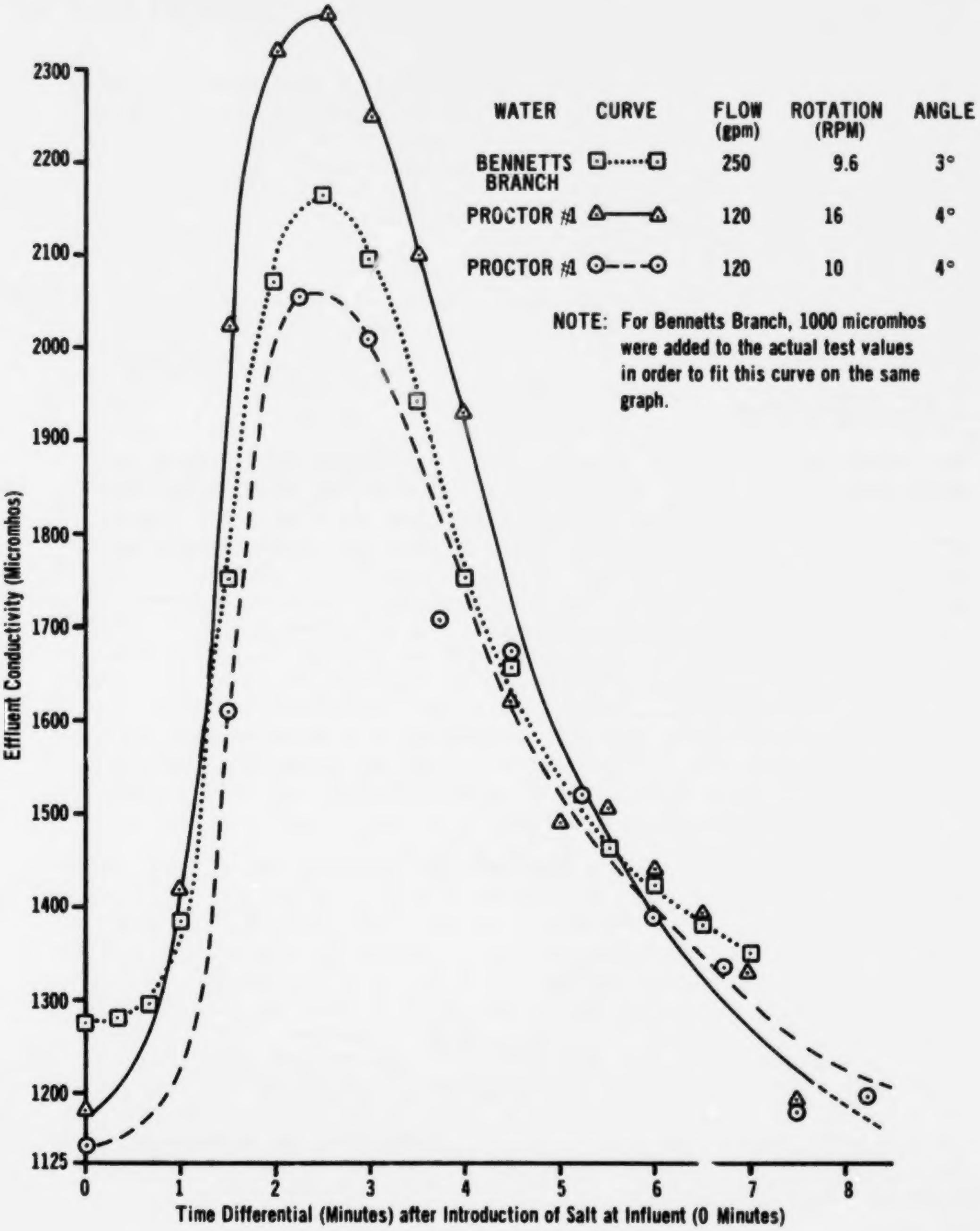
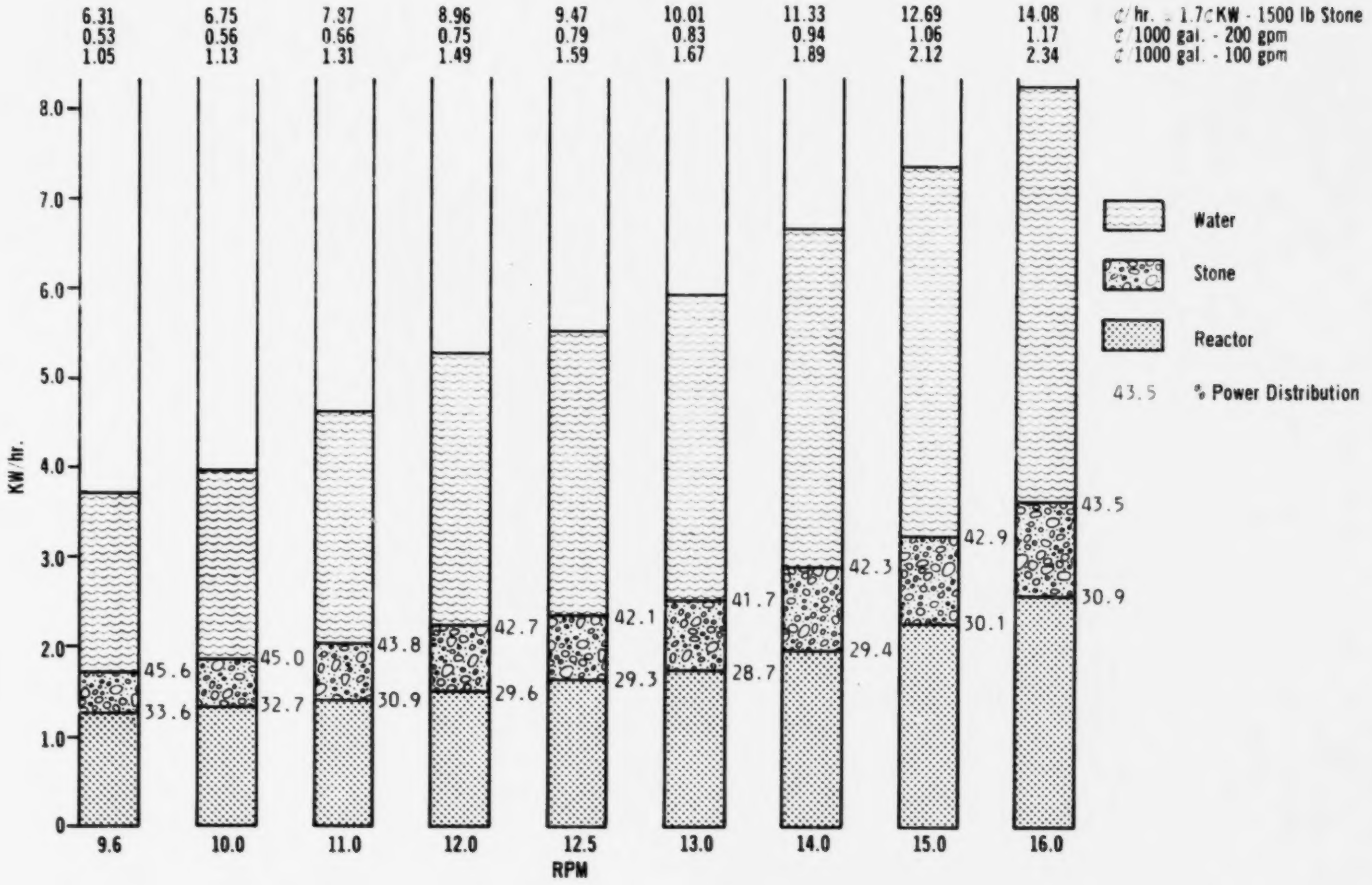


Figure 16. Distribution of power requirements for limestone reactor



The effective capacity of the rotary reactor is related to achievement of each of the necessary functions in response to the water and stone characteristics. Most critical is the reactor capability to provide adequate reactive stone surface which is a function of reactor stone load, of stone size distribution, of stone physical attrition, of stone distribution within the reactor, and of the stone consumption. The reactor must operate with excess stone present, but at an optimum and constant load. The power capability of this reactor will carry a load in excess of 1.5 tons of stone. Since power consumption varies with stone load, the optimum load must be limited to process requirements. The stone feed rate must balance the required stone consumption and must relate to the stone feeding system. Stone consumption results from reaction and the discharge of unreacted fines, with the latter determining reagent utilization efficiency.

With the introduction of a maximum load of stone to the reactor, the stone will distribute itself rapidly (within minutes) throughout the reactor length, governed by movement of different size particles through the three screens. The stone tends to move with the water flow toward the effluent end. Reactor slope control assists in establishing this rate; but appears to be effective only at angles less than 3° from the horizontal. Other configurations were not available for study as previously described.

The potential stone distribution within the reactor has extremes ranging from complete loading in section 1 (with + three-inch stone) to complete loading in section 4 (with - one-inch stone). The reactor design did not provide for exclusive use of pulverized stone due to the very low retention time (less than three minutes) which would result in excessive losses of reagent. In the first condition, attrition and comminution will be maximal in the first compartment, but stone surface area will be minimal, with three-fourths of the reactor being ineffective. With rotation time, the rapid attrition of the sharp edges of the large stone permit particle transfer through the screens to subsequent reactor compartments with the particles assuming smooth surfaces and round edges. A gradual, even distribution of stone throughout the reactor is approached. Larger total surface area of the stone is then achieved. However, the attrition-comminution rate decreases as the initial stones abrade to smooth surfaces and redistribute themselves to a reduced volume level throughout the reactor. With batch loading, it is during this initial period that conditions permit the greatest loss of unreacted stone. The attrition rate is also a function of particle size since the mass of larger particles falling will offer the greater impact for breakage. Thus, with increasing distribution and decreasing compartment loading in the reactor, the loss of unused stone decreases. Subject to the water flow and acidity loading, treatment will remain satisfactory with increasing stone surface area until the steadily decreasing attrition-comminution rate reaches a level that the total available reagent surface becomes inadequate. The production of a reactor-stone consumption level in excess of three pounds/minute was attainable. This would meet the requirements of CMD water having 2,000 ppm acidity at a flow rate of 300 gpm. This rate is beyond the design and hydraulic capacity of the reactor with the existing end-feed design.

The other loading extreme (minus one-inch stone which moves rapidly to reactor section four) would also limit reactor volume use and, due to greatly reduced attrition, would limit production of reactive stone surface despite large surface area. The proximity of this section to the effluent port would provide functional retention time of seconds, leading to inadequate reaction and excessive stone loss.

The optimum stone loading operation would require a uniform stone distribution throughout the reactor, controlled size distribution in each reactor section, and a continuous, uniform stone feed rate of specified size distribution. Ideally, the feed rate would precisely conform to the theoretical reagent requirements of the feed water. These conditions will vary with stone properties, especially hardness. Design of a reliable reactor feed system is feasible and would limit labor requirements to filling the stone feed hopper and maintenance.

Stone was introduced in the rotary reactor via a vibrating feeder from an 1,800 lb hopper. The hopper was loaded by a mobile conveyor belt. The stone feed rate to the reactor was adjustable by varying the amount of vibration and the size of the hopper diaphram. The reliability of feed control varied extensively with stone size distribution and moisture content. Mixed sizes (- three-inch) were least reliable, while narrow size ranges (especially 1/4- to one-inch) were most satisfactory. Limited moisture content of the stone tended to resist flow while operations during rain gave enhanced flow. Some difficulties were experienced from freezing during intermittant operation in cold weather. The stone feed equipment design is optimum for a given product with a limited range of feed rates, and will have decreased reliability with deviation from these specifications. The wide range of stone feeds and rates considered increased the control difficulties.

The Hollywood unit has been applied satisfactorily to CMD volumes in excess of 200 gpm, with higher levels limited by the reactor feed design. Evidence suggests a capacity of at least 400 gpm is entirely feasible for this size of vessel. It would be recommended that treatment capacity be increased by enlarging the drum diameter and reaction response by increasing drum length.

The economic, operating, and design criteria basic to the selection of the limestone treatment in contrast to other reagents for a given water is most complex. Figure 17 illustrates the alternate routes for the limestone-water reaction.

The more significant design parameters for limestone treatment given by unit operation follow:

Reagent Control

1. Consumption rate - lbs/hr.
2. Specifications for limestone grade, size, and hardness.
3. Type of storage (bins or piles).
4. Feeders and hopper design.
5. All weather provisions.

Limestone Neutralization Unit

1. Mode of operation (mixer or rotary mill).
2. Working volume and excess capacity allowance.
3. Weight and volume of limestone charge.
4. Pulp retention time vs. ferrous iron content.
5. Rotation or mixing requirements.
6. Gas exchange capability (oxidation and CO_2 removal).
7. Supplementary reagents required, e.g. $Ca(OH)_2$.
8. Materials of construction.
9. Cold weather provisions.

Iron Oxidation Unit (CO_2 Elimination)

1. Type (tank or pond).
2. Function - pre- or post-neutralization oxidation, aeration or bacterial.
3. Aeration capacity required by iron II concentration.
4. Retention time required.

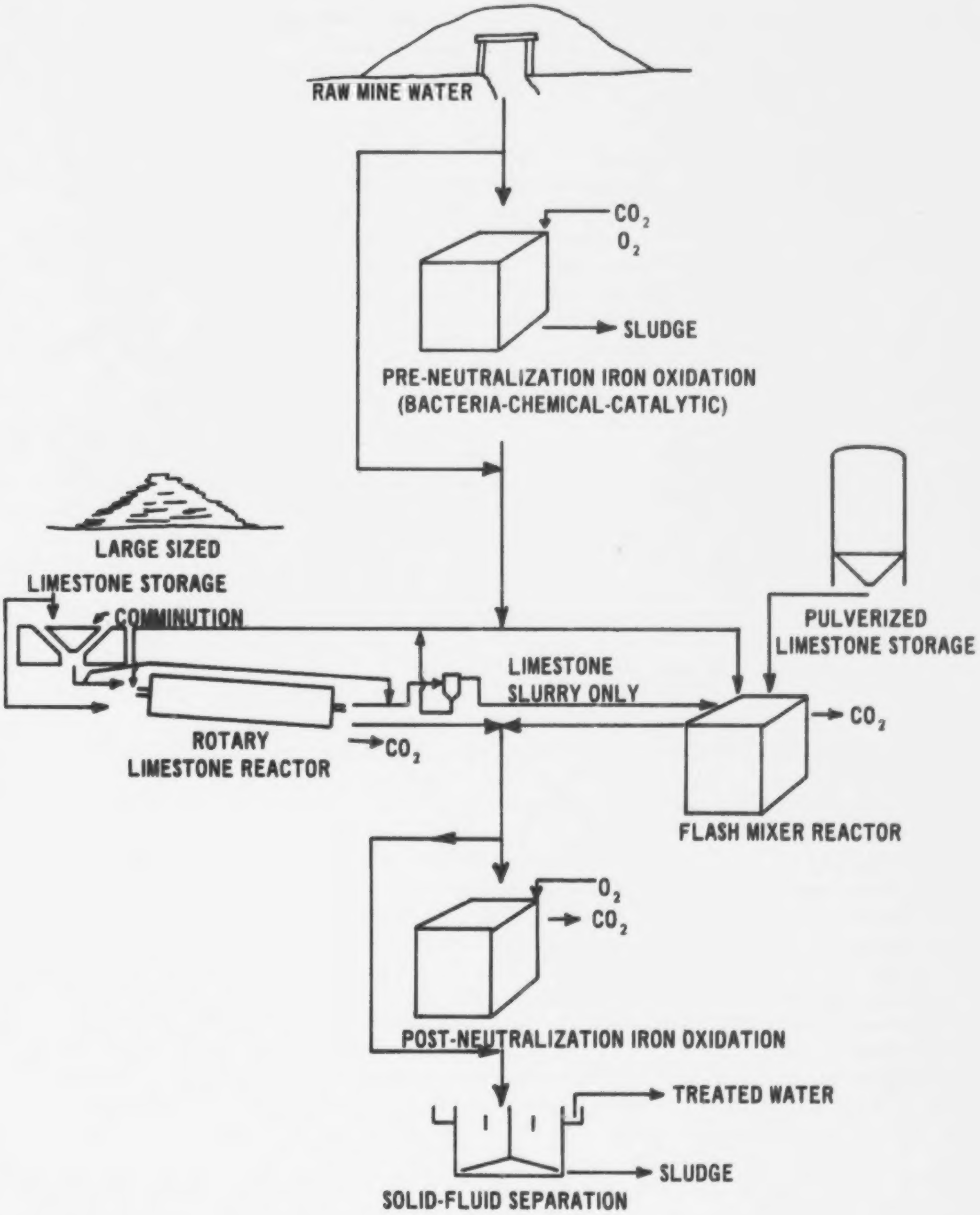
Settling Lagoon(s)

1. Settling rate of solids.
2. Loading rate (feed flow rate and percent solids).
3. Range of allowable rising velocities (gpm/ft^2).
4. Sludge bulk density.
5. Sludge storage capacity.
6. Provision for sludge removal.
7. Baffles.
8. Multi-pond systems.
9. Cold-weather provisions.

Thickener

1. Settling rate of feed slurry (including recycle).
2. Optimum feed solids content.

Figure 17. Alternate unit operations for limestone neutralization


RAW MINE WATER
CO$_2$
O$_2$
SLUDGE
PRE-NEUTRALIZATION IRON OXIDATION
(BACTERIA-CHEMICAL-CATALYTIC)
LARGE SIZED
LIMESTONE STORAGE
COMMINUTION
PULVERIZED
LIMESTONE STORAGE
LIMESTONE
SLURRY ONLY
CO$_2$
ROTARY
LIMESTONE REACTOR
CO$_2$
FLASH MIXER REACTOR
O$_2$
CO$_2$
POST-NEUTRALIZATION IRON OXIDATION
TREATED WATER
SLUDGE
SOLID-FLUID SEPARATION

3. Maximum underflow solids contnet.
4. Allowable rising velocity range - gpm/ft^2.
5. Ratio of recycle volume.
6. Operating characteristics - w/wo flocculant.
7. Feed flow rate.
8. Temperature effects.

Filter
1. Solids filtration rate - lbs/24 hrs/ft^2 filter area.
2. Liquid filtration rate - gpm/ft^2 filter area.
3. Type filter, filter area, filter medium.
4. Feed load - lbs solids/24 hrs/1,000 gallons CMD/ppm acidity.
5. Filter aid (pre-coat) requirement.
6. Flocculant requirement - lbs/1,000 gallons feed slurry.
7. Percent solids in filter feed.
8. Percent solids in filter cake.
9. Materials of construction.
10. Vacuum requirements.

Sludge Drying Basin
1. Bed design including porosity and drainage rates, area and depth.
2. Percent solids in feed slurry and "dried" solids.
3. Settling rate of feed slurry.
4. Mass of "dried" solids per 1,000 gallons feed water.
5. Precipitation and evaporation rate of the immediate area.

ROTARY LIMESTONE REACTOR DATA

The capability of the limestone neutralization process was demonstrated under severe conditions utilizing highly ferrous Proctor No. 2 waters. Water quality specifications were attained. Reasonable reagent use efficiency was experienced. Any coal mine drainage of lesser pollution loading will also respond satisfactorily. The iron II content was biochemically oxidized prior to neutralization and the waters were neutralized in the limestone reactor. The overall flowsheet is detailed in Figure 18 and the supporting conditions and analyses given in Table 18. The bacteria inoculation is used only to initiate the biochemical oxidation system and need not be maintained. The weight of limestone was evaluated volumetrically. The weight of given bulk volume of reagent was established. Stone was added through a vibrating feeder with nearly constant manual attention. The suspended solids in the effluent of the limestone reactor for this test were higher than normal, the value usually being closer 2,000/mg/l.

The limestone reactor sludge solids settled rapidly, with one-third final sludge volume being achieved within three minutes and nearly

Figure 18. Biochemical iron oxidation - limestone neutralization system

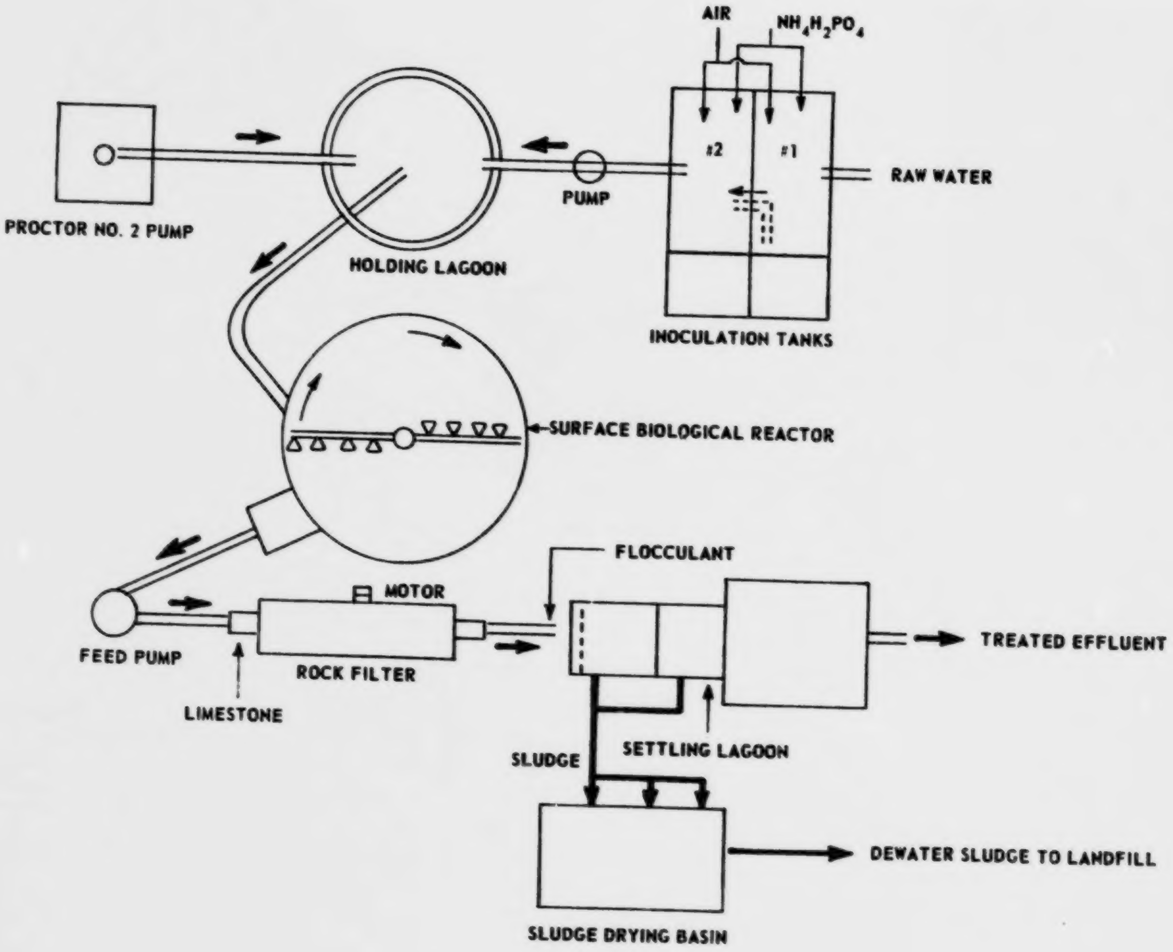

AIR
$NH_4H_2PO_4$
#2
#1
RAW WATER
PROCTOR NO. 2 PUMP
PUMP
HOLDING LAGOON
INOCULATION TANKS
SURFACE BIOLOGICAL REACTOR
FLOCCULANT
MOTOR
FEED PUMP
ROCK FILTER
LIMESTONE
TREATED EFFLUENT
SLUDGE
SETTLING LAGOON
DEWATER SLUDGE TO LANDFILL
SLUDGE DRYING BASIN

Figure 18 (continued). Schematic drawing og biological iron-oxidation limestone neutralization system

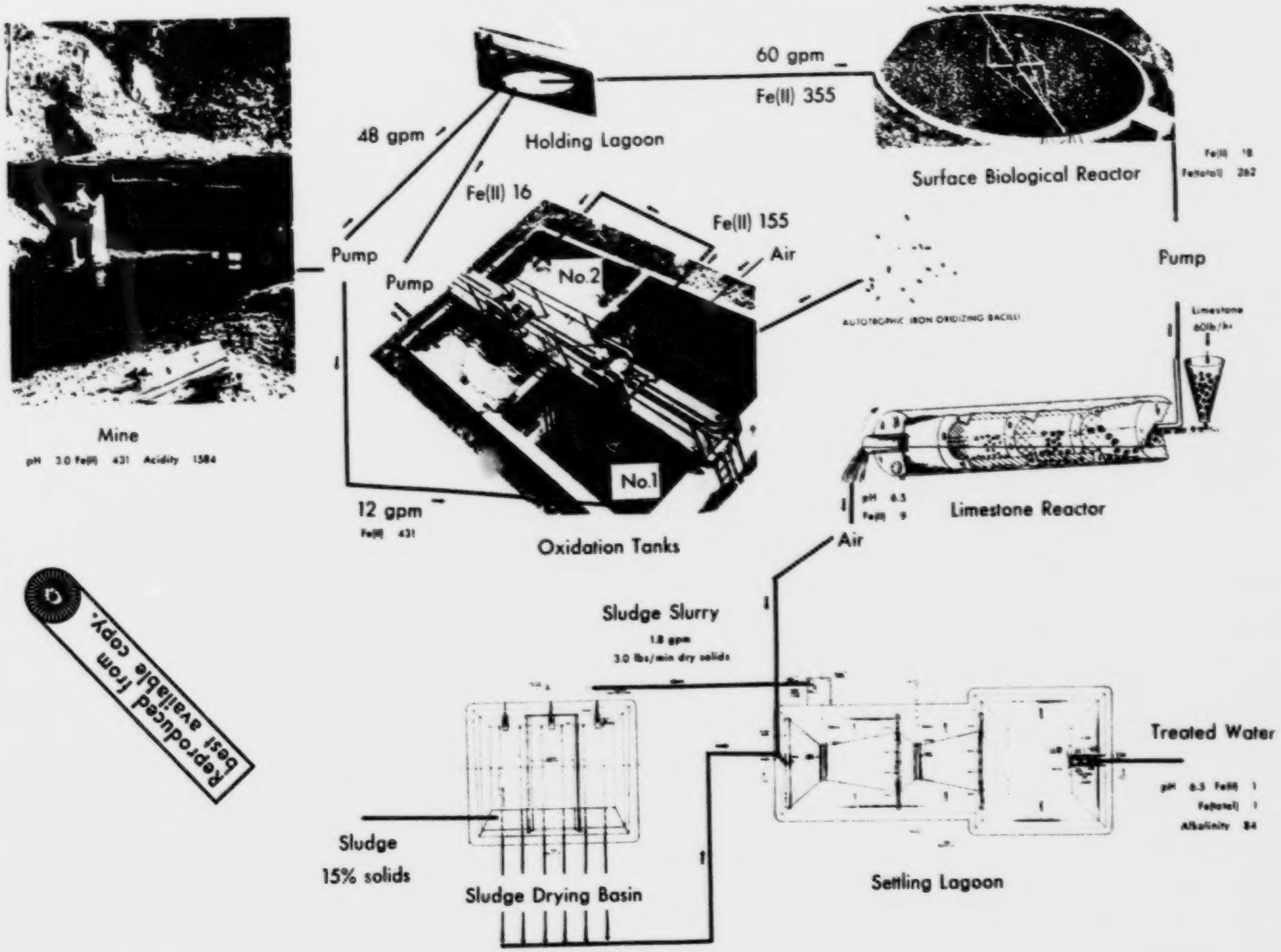

Mine
pH 3.0 Fe(II) 431 Acidity 1584
Pump
48 gpm
Holding Lagoon
60 gpm
Fe(II) 355
Surface Biological Reactor
Fe(II) 18
Fe(total) 262
Pump
Limestone
60lb/hr
Pump
Fe(II) 16
Fe(II) 155
Air
No.2
No.1
AUTOTROPHIC IRON OXIDIZING BACILLI
12 gpm
Fe(II) 431
Oxidation Tanks
pH 6.5
Fe(II) 9
Air
Limestone Reactor
Sludge Slurry
1.8 gpm
3.0 lbs/min dry solids
Treated Water
pH 6.5 Fe(II) 1
Fe(total) 1
Alkalinity 84
Sludge
15% solids
Sludge Drying Basin
Settling Lagoon

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Table 18. NEUTRALIZATION OF CMD BY LIMESTONE IN THE ROTARY REACTOR

Neutralization of CMD by Limestone in the Rotary Reactor
Test Period 9/2-10/71

Reactor Conditions

Speed - 13 rpm; Power consumed - 7.88 kWH/Hr. @ 19-23 amperes; Slope - 4°

Limestone Reagent

Seam - Valentine (south western end)

Location - Oak Hall, Centre County, Pa.
Neidigh Bros. Limestone Co.

Cost
Stone $1.33/ton
Transportation 1.95/ton (65 mi.@ 3¢/ton/mi.)

Chemical Analysis

Moisture = 2.0%

Dry Basis:
CaO = 50.0%
$CaCO_3$ = 85.0%
$MgCO_3$ = 1.5%
R_2O_3 = 0.5%

Hardness

Los Angeles Rattler Test - B Grading
100 revolutions - 5.6% passing
500 revolutions - 25.7% passing

Feed Rate - 60 lbs./hr
continuous feed

10% excess theoretical

Size Analysis:

Screen Size		Direct	Cumulative Wt.%	
Passed	Retained	Wt.%	Retained	Passing
	1/2 inch	39.0	39.0	100.0
1/2 inch	3/8 inch	10.0	49.0	61.0
3/8 inch	8 mesh	29.5	78.5	51.0
8 mesh	60 mesh	16.3	94.8	21.5
60 mesh	200 mesh	2.7	97.5	5.2
200 mesh		2.5	100.0	2.5

Process Flow Analyses

Source - Proctor No. 2
Flow - 60 gallons/minute
Process Temperature - 18°C
Theoretical limestone requirements - 0.0127 lb./gal.
Theoretical requirement of stone reagent utilized - 0.0152 lb./gal. or 0.90 lb./minute.
Flocculant addition - 5 ppm Atlasept 1A1 as 0.2% wt. solution.

	pH	Acidity mg/1 $CaCO_3$	Alkalinity mg/1 $CaCO_3$	Fe II mg/1	Fe III mg/1	Al mg/1	SO_4 mg/1	Cond. μ mhos.	Sus. Solids mg/1
Feed Water	2.8-3.2	1,520[c]		374	56	230	2346	2490	
Limestone Reactor									
Influent	2.9-3.2	1.500		26	183				
Effluent	6.2-6.6		128	20	35				
Settled Solids									4,232[a] 19.6% wt./vol.
Settling Lagoon									
Influent	6.40			6	1				
Effluent	6.48		78	<1	<4				
Settled Solids									9.5% wt./vol.[b]

Sludge Control:

Assuming transfer of about 2,600 gallons of sludge at 9.5% wt./vol. solids daily from settling lagoon to sludge drying basin over period of one hour, would add about 3 inches of sludge depth (assuming 1400 ft^2 area in a compartment). With the three compartment sludge drying basin, the transfer cycle would be: daily transfer of three inches of sludge for two consecutive days to each compartment.

Day No.:	1	2	3	4	5	6	7	8	9	10	11	12
Compartment No.:	1	1	2	2	3	3	1	1	2	2	3	3

Experience shows this would result in 1.50 inches/6 days of sludge level at solids content between 15 and 45 wt. % solids depending on drainage rate and weather: i.e., upon dewatering the settled solids reduced to about 1/4 their original volume. This sludge accumulation rate would require removal of sludge from the drying basin about every 144 days representing about 600 yds.^{3} or 450 ton (about 110 lb./ft^3). The sludge drying basin was emptied with a 2 yd.^{3} bucket on a model 922 caterpiller unit on cats. Two trucks with a total capacity of 15 ton carried the dewatered sludge about 1/2 mile for landfill. The sludge drying and volume reduction continues, reaching a final volume between 25-33 percent of the dewatered sludge and a water content related to the ambient conditions (1-10%).

---

[a] 0.035 lb. dry sludge/gallon or 2.12 lb./min.
Settled sludge volume - 3.0 vol. % in 10 minutes.
[b] 0.03 gal. sludge/gal. process influent or 1.8 gal./minute.

Composite sample during sludge transfer, value represents dilution with transfer water. Samples taken from lagoon bottom have exceeded 30 wt. %.

[c] Modified procedure - (ph 7.3) - 1,433 mg./1.

all of it within ten minutes, without the use of flocculants. These limestone-produced sludges (Type III settling behavior) do not exhibit a clear settling interface and their settling rates can best be observed by measuring sludge build-up. The supernatant water tends to be cloudy, but its clarification rate and level can be enhanced by the use of high molecular weight flocculants. A very high flocculant dosage was employed during this test, seeking indications of sludge balling with carbonate sludges, as was attained with hydroxide sludge. Positive results were attained which suggests that a bottom feed-type of solid-fluid separation operation would be especially effective for this type of sludge. In such an approach, the settling sludge blanket serves to collect and filter the flocculating colloidal particles resulting in an extremely clear effluent. The bottom fed weir (see Figure 11, Appendix B) to the settling lagoon provided sufficient retention time to settle the flocculated sludge and allowed a very clear effluent to enter the settling lagoon. This operation was effective to the degree that, had a proper pump been available to remove the settled sludge from the bottom of the weir, the settling lagoon would have been completely superfluous. Such a procedure would have major advantages in both capital and operating costs.

Further details of the overall limestone process from the Hollywood Facility have been presented elsewhere[98, 99]. The response to other limestones was satisfactory but it varied according to particle size, purity and hardness as described previously in this report. Dolomite also responded satisfactorily but was more difficult to control having reduced reactivity and greater hardness thus limiting maintenance of adequate available reactive surface. The use of pulverized limestone (-200 mesh rock dust) was helpful as a supplemental feed to the rotary limestone reactor to complete the reactions and to add excess alkalinity to the effluent. However, beyond requiring a separate feeding system, control of its addition must be exceptionally tight to prevent use of excess reagent.

The use of pulverized limestone in the flash mixer and Densator are not detailed here. These units were not designed for this application and did not provide sufficient agitation velocity to maintain stone suspension. Although a satisfactory effluent could be maintained, the resulting sanding of the unreacted reagent required the introduction of excessive quantities of pulverized limestone.

PROCESS COST ANALYSIS

A process cost analysis is given in Table 19. The total cost, as calculated, is based on charges experienced at the Hollywood Facility and includes all phases of the treatment process except taxes

Table 19. ESTIMATION OF COSTS - LIMESTONE TREATMENT PROCESS
(basis: cents/1,000 gallons)
(process conditions: table)

Unit Operation	Power[h]	Reagents	Labor[b]	Maintenance	Amortization[a]	Inoculation[c]	Contract[m]	Totals	%	Totals Excluding Amortization	%
Water Collection	3.25			0.19[d]	5.01			8.45	11.12	3.44	14.55
%	38.5			2.2	59.3						
Transfer - Plant					1.56			1.56	2.05		
%					100						
Holding Lagoon					4.46	0.58		5.04	6.63	0.58	2.45
%					88.5	11.5					
Iron Oxidation		1.67[e]	0.95[f]		9.10	0.58		12.30	16.19	3.20	13.54
%		13.6	7.7		74.0	4.7					
Pump to Reactor	0.56							0.56	0.74	0.56	2.37
%	100.0										
Neutralization	3.72	2.74[g]	0.87[i]	1.29[j]	11.92			20.54	27.04	8.62	34.46
%	18.1	12.4	4.2	6.3	58.0						
Solid-Fluid Sep.		1.25[k]	3.47[l]	0.10	13.67			18.49	24.34	4.82	20.39
%		6.8	18.8	0.5	73.9						
Sludge Dewater.					6.61		2.42[n]	9.03	11.89	2.42	10.24
%					73.2		26.8				
Totals	7.53	5.66[o]	5.29	1.58	52.33	1.16	2.42	75.97	100.00	23.64	100.00
%	9.91	7.45	6.96	2.08	68.88	1.53	3.19	100.00			
Totals (excl. amor)	7.53	5.66[p]	5.29	1.58		1.16	2.42	23.64			
%	31.85	23.94	22.38	6.68		4.91	10.24	100.00			

NOTES:

[a] Amortization based on direct cost for ten year life.
[b] Labor based at $3.00/hour.
[c] Assumes $1,000.00 equipment cost, $100.00/year supplies and $165.00/year labor.
[d] Lubrication 15 minutes/day - 0.13, replacement parts 0.06.
[e] 5 ppm $NH_4H_2PO_4$ @ 0.10¢/lb. - 0.40; filter cleaning - HCl - 1.27.
[f] 100 man hours/ year.
[g] 16.7 lb/1,000 gal. @ 0.164¢/lb.
[h] @ 1.7¢/KWHr.
[i] Fill limestone hopper - 15 minutes/day.
[j] Lubrication 15 minutes/day - 0.13; replacement parts - 1.16.
[k] Flocculant - 1 ppm @ $1.50/lb.
[l] Prepare flocculant - 10 minutes/day - 0.56; sludge transfer - 1 hr/day - 2.91.
[m] Remove sludge from drying basin.
[n] Charges for caterpiller, two trucks and labor - $0.67/Ton.
[o] Alkali costs only 3.6%.
[p] Alkali costs only 11.6%.

and administration overhead. The values listed were derived directly from pilot plant costs and cannot be realistically related to industrial operation. Obviously, they are high. The costs are also shown without the inclusion of amortization values. This compilation offers realistic expression of the factors which must be included in cost considerations. The low percentage of total costs related to the reagent is noteworthy.

Reasons for High Costs

Most of the equipment was not being utilized at optimum capacity. The equipment would permit handling a flow of at least 300 gpm which would reduce amortization charges by 80 percent. The existing conditions of the limestone reactor and particular limestone reagent were the limiting process flow factors. The surface biological reactor was shown to perform satisfactorily with these waters at levels at least two times those utilized in this test. Increased flows would incur no additional labor costs and very little increased power costs. The reagent and sludge control costs are closely proportional to flow levels. The amortization costs are very high since they involve research equipment designs. A ten-year depreciation period was employed. The equipment can be expected to provide useful service for a longer period.

SECTION XIII

BIOCHEMICAL OXIDATION OF IRON II

The use of limestone to neutralize coal mine drainage has several limitations. Limestone has a slowerreaction rate which is controlled by its limited solubility. The limestone particles are subject to surface coating with reaction products. Further, the maximum pH which can be developed in the aqueous system with limestone cannot exceed about 8.3 due to several chemical equilibria, including the presence of soluble carbon dioxide. In fact, only under extremely vigorous agitation and with the use of very finely-divided limestone particles can the pH of the treated effluent rise above 7.0. Although this situation prevents overtreatment of CMD, the lower pH levels place restrictions upon oxidation of ferrous iron, which normally follows neutralization and is necessary to complete the treatment[42, 47]. The practical limitations of limestone found in these evaluations have also been cited in Travers[48, 49] and by Birch[50, 51] who combined lime with limestone to meet water specifications.

As described in the limestone neutralization section (Table 16), our experience indicates a practical limitation of the use of this desirable reagent to waters containing less than about 100 mg/l iron II. This arbitrary value appears to be related to the oxygen solubility attainable. Waters with somewhat higher levels of iron II may be treated satisfactorily if the limestone neutralization reaction is followed by lengthy air oxidation periods in the presence of large excesses of pulverized limestone[52]. During this oxidation period the limestone must be maintained in suspension.

To permit the effective use of limestone neutralization in ferrous waters above the 100 mg/l level, it becomes desirable to oxidize iron II in an acid system. Oxidation by air (oxygen) under these lower pH conditions is impractically slow. Stronger chemical oxidants such as peroxides, permanganate, and chlorine may be used but are deemed too expensive--at least to the degree that their use would increase the costs to the level which would preclude the advantages of limestone.

It is generally accepted that autotrophic or chemolithotrophic bacteria, such as *Ferrobacillus ferroxidans* and *Ferrobacillus thiooxidans*, play a role in the development of mineralized mine drainage from pyrites[53]. It is also known that these bacteria will oxidize soluble ferrous iron in an acid solution[54]. This response by the bacteria may be considered a catalysis of the reaction between atmospheric oxygen and iron II ions, which is thermodynamicallyfavored, but apparently kinetically-hindered. Glover has reported utilizing this biochemical iron oxidation as part of a total coal mine drainage treatment process in a small pilot plant[40, 41, 42]. Huddleston and Whitesell[55, 56] have reported laboratory studies of this oxidation reaction. Silvermann[57], Lundgren[58], and Unz[59], among others, studied the nature of these bacteria.

The Hollywood Facility was designed to permit the biochemical oxidation of iron II. In addition to the vat-type of oxidation vessel, somewhat similar to Glover's design but simpler and with surface aerator, Lovell included a "trickling filter"-type of unit operation at Hollywood to carry out this biochemical oxidation. Bench scale studies made in 1966 established the feasibility of this latter concept, including its combination with a rotary kiln-type limestone reactor.

This "trickling filter-type" of unit operation, termed a "surface biochemical reactor," has several inherent advantages: 1) No energy is required to control the water distribution or to introduce oxygen and carbon dioxide into the reacting water. The pressure supplying the water to the reactor, whether achieved by pump or gravity pressure head, should be adequate to distribute the water. Porosity through the media bed permits distribution of oxygen and carbon dioxide from the air. 2) The reactor provides reaction surface area which may be controlled by selection of the media particle configuration. 3) Capital cost of the unit is low. Designs less complex and costly than the Hollywood reactor may be feasible based on this experience. Glover[52] cited as a disadvantage of the "percolating filter" concept a tendency to build up insoluble iron oxidation products. Although such deposition occurs, it is the high surface area associated with the deposition which supports the bacteria and provides interphase transfer of nutrients. The cementitious nature of the deposit which builds on the surface of the media in the reactor is indicated in Figure 19. This reaction area creates the superior oxidation rates (300 to 400-fold higher) reported in this study. The build-up in the surface reactor at the Hollywood Facility did not reach a level, in over one year's operation, detrimental to flow through the filter. Continual deposit growth will lead to channelling and a reduction in efficiency. Preliminary tests indicate several approaches to remove the sludge: partial solubilization and water scrubbing *in situ*, or after drying, the deposit will spall from the media surface with minor vibration. Chemical control to prevent iron III hydrolysis is possible but does not appear to be feasible as a solution to the deposition problem.

The two controlling factors in this biochemical process include the growth rate of the bacteria and the rate at which the bacterial

Figure 19. Deposit on Surface Biochemical Reactor

system oxidizes the ferrous ion. In a continuous treatment system, the retention time in the reactor is the critical design factor.

Parameters which establish the bacterial growth rate include: inoculation level, temperature, iron II concentration, bacterial nutritional requirements (as oxygen, carbon dioxide, phosphate, calcium, nitrogen, and trace elements), agitation, and the unique characteristics of a given bacterial strain, including tolerable population levels. These parameters determine the required bacterial population. Evidence suggests minimal cell counts of 5×10^6 cells/ml are necessary, whereas higher levels are more desirable. Cell levels in excess of 10^7 cells/ml have been maintained in continuous flow systems, whereas counts approaching 10^9 cells/ml have been achieved in laboratory cultures. Glover[52] reports the "activated sludge" serves to maintain an appropriate concentration of bacteria--a conclusion not wholly supported by the current studies.

The bacterial iron oxidation rate is dependent upon maintenance of a viable cell population (growth rate, cell viability time, system cell loss), temperature, iron II concentration, pH, and at least one inadequately-defined parameter which is related to liquid-solid interface surface areas. This latter factor(s) may be involved in Glover's "activated sludge", and may explain the high level of oxidation activity at the surface of natural stream "yellowboy". Glover reports the bacterial sludge to have a low cell count, which was confirmed in these studies. It is the liquid-solid interface area which is believed to be responsible for the much higher oxidation rate found in the surface biochemical reactor during these studies in contrast to those reported in vat-type systems[56].

Associated with the effect of pH upon the biochemical oxidation of iron is another related equilibrium: the hydrolysis of the ferric ionic species (presumably as some sulfate complex) which results in the insoluble hydrous ferric oxide deposit. When the iron II oxidation is carried out chemically with oxygen at higher pH levels, this hydrolysis proceeds very rapidly, reducing the soluble iron species to less than one mg/l. As the oxidation is carried out at low pH levels (pH 2.9-3.0), the hydrolysis reaction is slow and the resulting iron III species tend to remain in solution. Under acid conditions, the biochemical oxidation system effluent remains relatively free of suspended iron oxide, but does deepen its amber-like color as the concentration of the ferric ion species increases. The sludge deposition appears to grow from nucleation sites rather than direct precipitation containing ferric ion. A slight increase in pH (by 0.1-0.3 units from addition of lime slurry) of the water slowly develops a colloidal "haze" but the formation of distinct particles is retarded. Higher pH levels from minor lime additions to waters are apparently detrimental to the bacteria metabolism. Other bacterial species may prefer higher pH levels. Glover's "activated sludge"[42] may play a role in this hydrolysis reaction, perhaps beyond that as a carrier of bacterial cells. The recycled "activated sludge" may serve as a physical growth nucleus favorable to this reaction. This phenomenon may be a key to the further enhancement of this biochemical oxidation process and may be the undefined parameter which is related to surface area. In these studies, the merit of separating and recycling the sludge formed after biochemical iron oxidation could not be supported, and was discontinued.

The chemical analysis of the sludge developed by biochemical oxidation is given in Table 20. It is characterized by high iron and sulfate concentrations as well as low aluminum concentrations, when contrasted to sludge from alkali neutralization (see Table 50). Apparently between 15 and 16 percent of the iron present is associated with the sulfate. The high sulfate content suggests another advantage of biochemical iron oxidation in achieving some sulfate concentration reduction. The potential to enhance this sulfate reduction is of great interest.

The iron oxidizing bacteria were named *Ferrobacillus ferrooxidans* by Leathen et al.[60] but are regarded by research workers as closely related to members of the Thiobacillus genus[61] since many of them can oxidize sulfide compounds as well as iron. These bacteria are variously referred to as *Ferrobacillus ferrooxidans*, *Ferrobacillus thiooxidans* and *Thiobacillus ferrooxidans*. In waters containing Iron II which is associated with sulfide, there no doubt exist several strains of these bacteria that vary slightly in appearance and in their biochemical activities.

Table 20. CHEMICAL ANALYSIS OF SLUDGE FROM BIOCHEMICAL OXIDATION OF COAL MINE DRAINAGE WEIGHT PERCENT

	Settled sludge from oxidation tank	Settled sludge from thickener
Fe	42.54	40.34
SO_4	17.39	15.68
Loss on ignition 180°C	14.16	15.44
P_2O_5	2.44	2.25
SiO_2	0.29	1.13
Ca	0.21	1.95
MG	0.12	0.24
Al	nil	0.98

Members of this group of microorganisms are nonspore forming, gram-negative rods varying from 1.0 to 2.0 microns in length and having polar flagella. Figures 20 and 21 show one such isolate, Strain Z, that has slightly curved cells. They are approximately 1.5 microns in length and have a single polar flagellum. Figure 20 results from optical magnification through a phase contrast microscope, while Figure 21 is an electron microphotograph at 128,000 magnification.

The organisms can be observed *in vivo* with a phase contrast microscope and appear to be straight or slightly curved rods, actively motile and varying in diameter as well as length.

The organisms derive their basic metabolic carbon, hydrogen, and oxygen from carbon dioxide, water, and air. They grow slowly compared to common heterotrophic gram-negative rods, having generation times of about five or six hours at their optimum temperature of 25 to 28°C. Temperatures higher than 35°C are harmful to most natural isolates. Glover[52] reports 40°C to be immediately fatal. The organisms grow at colder temperatures, although their growth rate decreases with temperature. Laboratory experiments with several strains have indicated that at 10°C, the bacteria require from 24 to 30 hours to double their

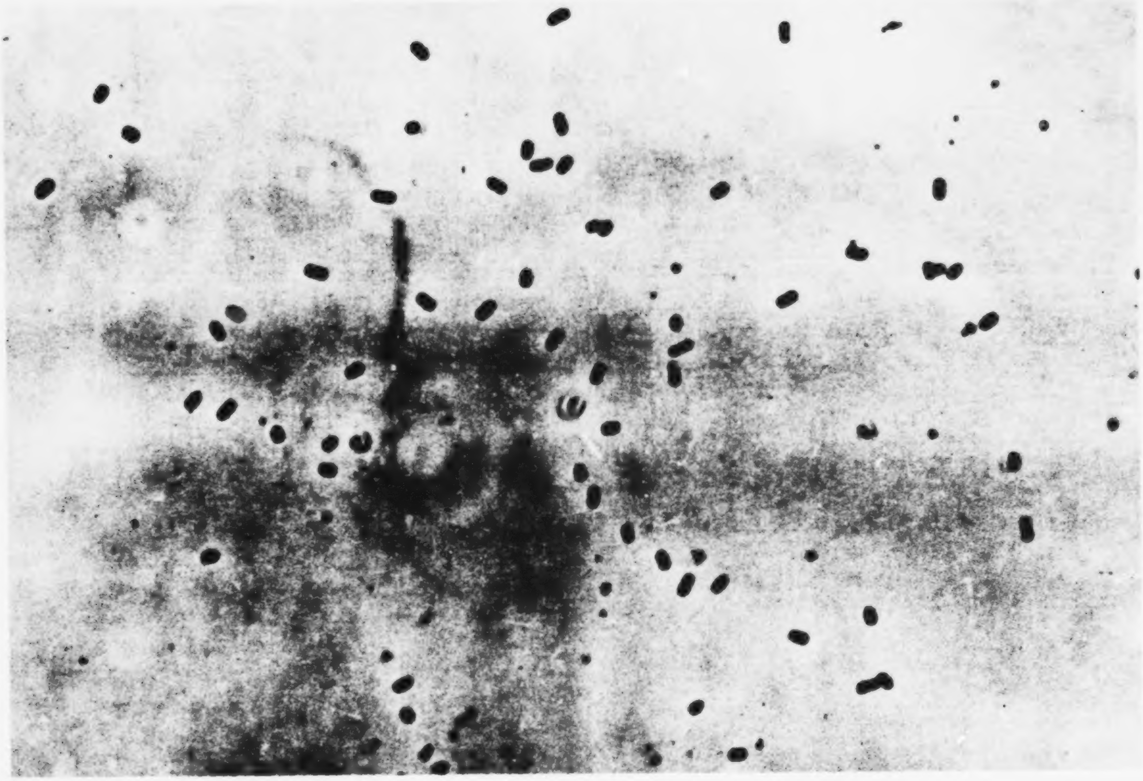


Figure 20. Microphotograph of Culture of Ferrobacillus Ferrooxidans Employed to Oxidize Iron II.

Figure 21. Electron Microphotograph of Ferrobacillus Cell

numbers. They achieve a low cell weight (1 g, dry weight) from the much larger amounts of iron (180 grams)[52].

A growth lag period of 10 to 20 days was reported by Glover[52]. High levels of inoculant encourages initial growth. If the initial plant system has cell levels of the order of 10^2 cells/ml, a much longer incubation period can be anticipated. The permissible pH range appears to be between 1.5 and 4.0, with an optimum near 3.0. The pH levels which might be specified in this biological process illustrate the complex nature of the reactions which occur. There are at least two optima which should be satisfied: the pH condition which relates to the growth levels of the bacteria, and that pH condition which relates to the Eh level most favorable to the iron II oxidation. The latter level increases with rising pH, thus developing conditions more conducive to chemical oxidation. Efforts to enhance the oxidation rates by maintaining higher pH levels in the biochemical oxidation tanks (by the addition of lime slurry) have not been advantageous--suggesting that the metabolic levels of the bacteria decrease in these higher pH ranges. This observation confirms a distinct role of the life process in this biochemical oxidation. Further study of the mechanisms involved is strongly indicated.

Under continuous flow conditions, Glover[42] reported iron II oxidation rates of nearly 11 mg/1/hr when maintaining an equilibrium concentration of 50 mg/1 iron II. Rates nearly three times greater were reported in ten-day-old laboratory cultures. Air was introduced by sparger into Glover's oxidation reactors. In subsequent, larger scale studies, inoculation of the water and slow development of an "activated" sludge, the rate increased to 120 mg/1/hr while maintaining a temperature of 15-20°C. The suspended solids were maintained near the one percent level. At near-freezing temperatures, the rate dropped to 12 mg/1/hr. The "activated" sludge was recycled and became very tacky and cohesive. He employed retention times of one hour at 20°C and up to five hours at a temperature near 0°C. Huddleston et al.[55] in 1969 reported much higher rates in laboratory cultures with synthetic solutions but were unable to sustain them in continuous flow with actual mine water (at 20-26.5°C) in a small pilot plant[56]. These authors found approximately 0.1 volume of air/volume of water to be optimum. They also utilized about 20 mg/1 each of ammonia and phosphoric acid as nutritional supplements.

PLANT OPERATIONAL DATA

Development of Bacterial Cultures

The bacteria employed were isolated from samples of Hollywood waters collected in sterile containers, and, in one instance (Strain Z), from a Central Pennsylvania coal refuse product derived from Lower Kittanning seam coal. Cultures were obtained by an enrichment technique employing either Silverman's 9K medium[57] or American Type Culture Collection Thiobacillus medium No. 65[62]. These media are highly selective for members of *Ferrobacillus* and *Thiobacillus*. The composition and preparation of the media is described in Table 21. During the culture enrichment, the media were inoculated at the ten percent levels, incubated at 26°C on a reciprocal shaker until the characteristic yellow-brown ferric precipitate developed in three to five days. The cultures were purified by tube dilution with the final tube being transferred to fresh, sterile media. Repeated transfers in the highly selective medium were assumed to result in a pure culture. Stock culture is maintained in cold storage for subsequent study and the development of inoculum. A sample of *Ferrobacillus ferrooxidans* ATCC 13661 was grown concurrently and exhibited behavior similar to cultures collected at the plant site.

Although the changing iron II levels during oxidation can be employed as evidence of the extent of bacterial action (and probably adequate as a control procedure in an operating plant), direct cell assays are helpful since cell population is obviously an oxidation rate parameter. Biological assays are too slow for control purposes, requiring from 48 hours to 14 days to yield a measurement. In addition, these species do not respond well to the plate-type counting assays commonly employed for coliform and other common types of bacteria. Huddleston et al.[55, 56] reported some cell population data. Direct cell counts are more practical.

Direct Microscopic Count Assays

Direct microscopic counts were made with a Petroff-Hauser bacterial counting chamber from observations with a phase contrast microscope. The microscope factor of this chamber was 2.0×10^7 per square. A minimum of 20 squares were evaluated. Significant results were limited to samples containing at least $1\text{-}5 \times 10^6$ cells/ml. Care was taken to pipette a representative sample of the suspension of microorganisms. The accompanying yellow precipitate caused some difficulty in filling the chamber and required care in observing the cells. In active cultures, the high motility of the cells made counting difficult. The addition of one-tenth volume of 95 percent ethanol served as a tranquilizer to slow cell action.

Tube Dilution Assays

Most probable number of cells was also established by a tube dilution procedure based upon Woodward's table of MPN[63]. A triplicate series of decimal dilutions comprised each test involving 10 to 10^{-4} ml for mine water samples and from 10^{-2} to 10^{-9} ml to estimate laboratory

Table 21. CULTURE MEDIA FOR IRON OXIDIZING BACTERIA

9K Medium

Salt Solution

$(NH_4)_2SO_4$	3.0 g
KCl	0.1 g
K_2HPO_4	0.5 g
$MgSO_4 \cdot 7H_2O$	0.5 g
$Ca(NO_3)_2$	0.01 g
$10N \cdot H_2SO_4$ to pH 2.8 approx.	1.0 ml
Distilled water	700 ml

Ferrous Sulfate Solution

$FeSO_4 \cdot 7H_2O$	44.2 g
Distilled water	300 ml
Adjusted pH 2.8 with H_2SO_4	

Autoclave solutions separately. When cool, decant ferrous sulfate solution from slight precipitate and combine aseptically with salt solution. (Silverman employed the medium at a pH 3.0-3.6; in these studies it was routinely adjusted to pH 2.8).

Medium No. 64

Solution A		Solution B	
$(NH_4)_2SO_4$	0.8 g	$FeSO_4 \cdot 7H_2O$	20.0 g
KH_2PO_4	0.4 g	Distilled water	200 ml
$MgSO_4 \cdot 7H_2O$	0.16 g	pH 2.8 with 1 N H_2SO_4	
Distilled water	800 ml		
pH 2.8 with 1 N H_2SO_4			

Autoclave separately, cool, decant as with Medium 9K and combine.

cultures. The dilutions were made using sterile, distilled water buffered with KH_2PO_4 (0.4 g/l) and H_2SO_4 to the pH range of 2.8-3.0. The culture tubes were incubated at 28°C for 10-14 days and were inspected daily. The formation of a yellow color and orange-brown precipitate was taken as the end point. The results with laboratory cultures using Medium No. 64 compared favorably with direct counts, usually being within a factor of two. However the results on natural mine water samples were less precise. The influence of other organisms in these waters may be responsible for this deviation. The tube dilution technique serves as a check for direct count measurements and provides a means of evaluating counts for samples containing less than 10^6 cells/ml. Cell counts on the four Hollywood water sources made in August 1969 were:

	Fe II mg/l	Cell/ml
Proctor No. 2 Borehole	600-650	90-250
Proctor No. 1	30-70	2-25
Bennett's Branch	20-25	40-240
Tyler Run	<5	1-50

Bacterial Growth Factors

Growth rates are definitely lower at pH 3.6 in comparison with the 2.8-3.0 range. No evidence has been developed of growth inhibition at increased oxygen levels. Flask agitation stimulates growth in comparison with stationary cultures. The addition of organic growth stimulants such as 0.005% yeast extract does not enhance and may even retard cell growth. The addition of calcium ion (as $CaCl_2$ or $CaCO_3$) in the range of 50-200 ppm to the No. 64 Medium stimulated growth of some of the isolated strains. This enhanced response was related to the calcium ion rather than the carbonate ion. The Hollywood waters all contain calcium concentrations in this range so that further calcium supplement should not be necessary. The response to additions of NH_4^+ and PO_4^{-3} ions is indicated in Table 22. Only Proctor No. 2 waters showed significant stimulation which suggests that the oxidation rate of highly ferrous waters may benefit from such additions. Phosphate alone was as effective as the ammonium ion or perhaps the combination of both.

Preparation of Inoculants

A 10 ml portion of stock culture, stored at 5°C, was utilized to inoculate 100 ml of No. 64 Medium in shaker flasks. Growth reached the 10^8 cells/ml in about 48 hours. A combined culture, totalling 500 ml from five shaker flasks, was used to inoculate a five gallon carboy of Medium No. 64 used to seed the tanks at the plant. The medium in the carboy was aerated by a simple glass sparger.

Initiation of the Biochemical System

Experience indicated the advantage of starting these systems from high cell count, pure, laboratory cultures rather than seeking to increase the population from a low cell count, natural water or attempt to introduce a "natural" culture from the surface of the yellowboy deposit in a stream. A series of batch steps is desirable in which a minimum of ten percent inoculant is utilized in each step. Typical batch oxidation rates are illustrated in Figure 22 while the conditions utilized are detailed in Table 23. In stage A (Figure 22), a 150 gallon tub was employed as the reactor with the matured growth being transferred to the sludge sump for stage B and subsequently to one of the concrete oxidation tanks (see Appendix B, Figure 13) for

Table 22. GROWTH SIMULATION OF IRON OXIDIZING BACTERIA
IN NATURAL MINE DRAINAGE
IRON II CONTENT (mg/l)

Water source	Incubation time			
	Initial	1 Day	2 Days	4 Days
Proctor No. 2				
Blank	665	-	529	<5
NH_4^+ 1 mg/l	665	-	520	<5
NH_4^+ 10 mg/l	658	-	474[1]	<5
PO_4^{-3} 10 mg/l	658	-	123[2]	<5
NH_4^{+1} 10 mg/l PO_4^{-3} 10 mg/l	652	-	208[3]	<5
	Original cell count		90 cells/ml	
	1 cell count		2,400 cells/ml	
	2 cell count		4,000 cells/ml	
	3 cell count		40,000 cells/ml	
Proctor No. 1	25	-	<5	
No observed change from control in two days.				
Bennett's Branch				
Blank	23.5	15.4	<5	
NH_4^{+1} 1 mg/l	23.5	15.4	<5	
NH_4^{+1} 10 mg/l	23.3	11.0	<5	
PO_4^{-3} 10 mg/l	23.3	8.8	<5	
NH_4^{+1} 10 mg/l PO_4^{-3} 10 mg/l	23.1	<	<5	

stage C. In proceeding to stage D, the third stage product was equally distributed between the two oxidation tanks which were then filled with raw wqter. A final iron II concentration of about 10 mg/l was reached within 24 hours. Little iron III is hydrolyzed and precipitated from solution during these batch oxidations. The amount of suspended solids near the end of stage A was about 150 mg/l which increased to about 300 mg/l in Oxidation Tank 1 toward the end of stage D. With the maturation of the culture representing two-10,000 gallon stages, the batch operation was converted to a dynamic multistage system by the continuous introduction of raw water.

In addition to the water loading rate to the surface reactor (>0.16 gpm/ft^2 surface), the retention time in the reactor is critical. The detailed response of this reactor, especially the media configuration, was not included in the present study but is being pursued. The retention time of the surface reactor was found to range between 4.7 and 5.3 minutes in the flow range considered. This parameter will not

Figure 22. Iron II oxidation rates with batch inoculation of mine drainage

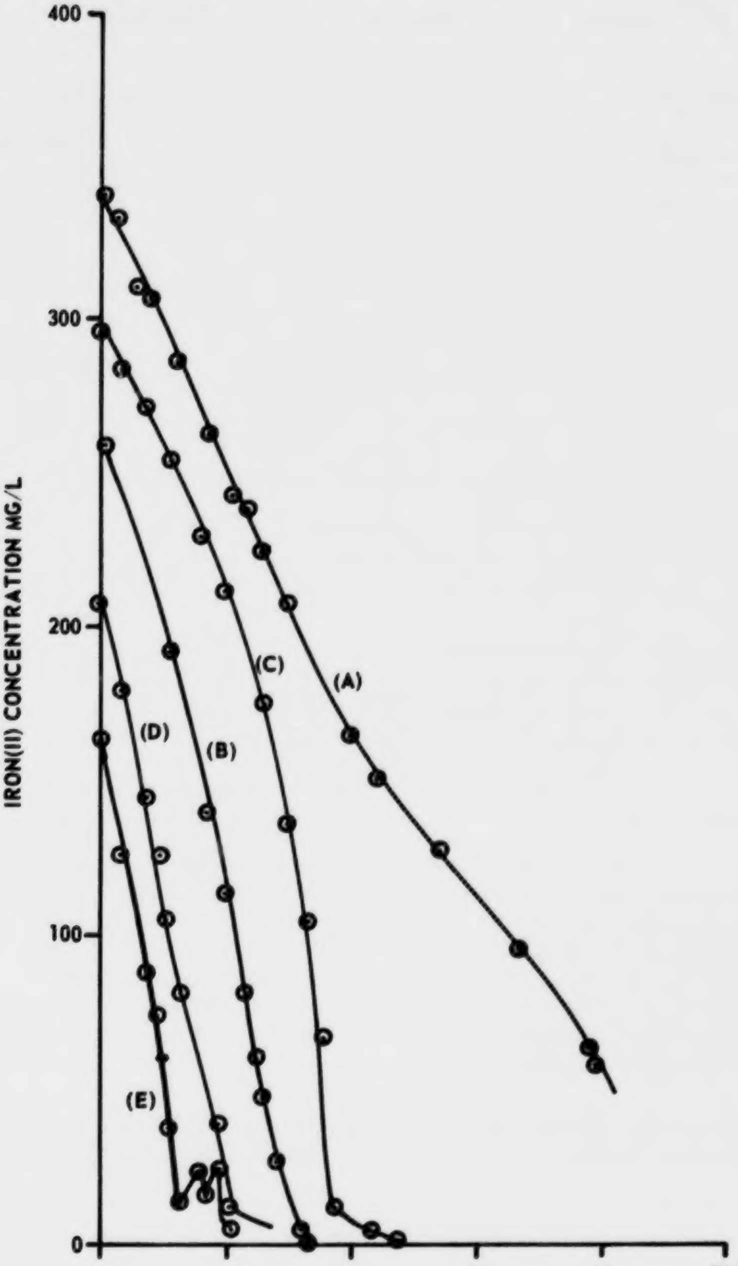


Table 23. DETAILED CONDITION OF BATCH INOCULATION

Stage	Reactor volume gallons	Inoculation level volume %	Overall oxidation rate mgFe/L/Hr.
A	150	3.2	3.6
B	1,257	10.8	7.9
C	9,930	12.4	7.6
D-1	9,930	50.0	9.8
D-2	9,930	50.0	11.5

		Stage A	Stage C
Nutrient supplement: (as $NH_4H_2PO_4$)	NH_4^+	1.6 mg/l	1.0 mg/l
	PO_4^{-3}	3.4 mg/l	5.1 mg/l

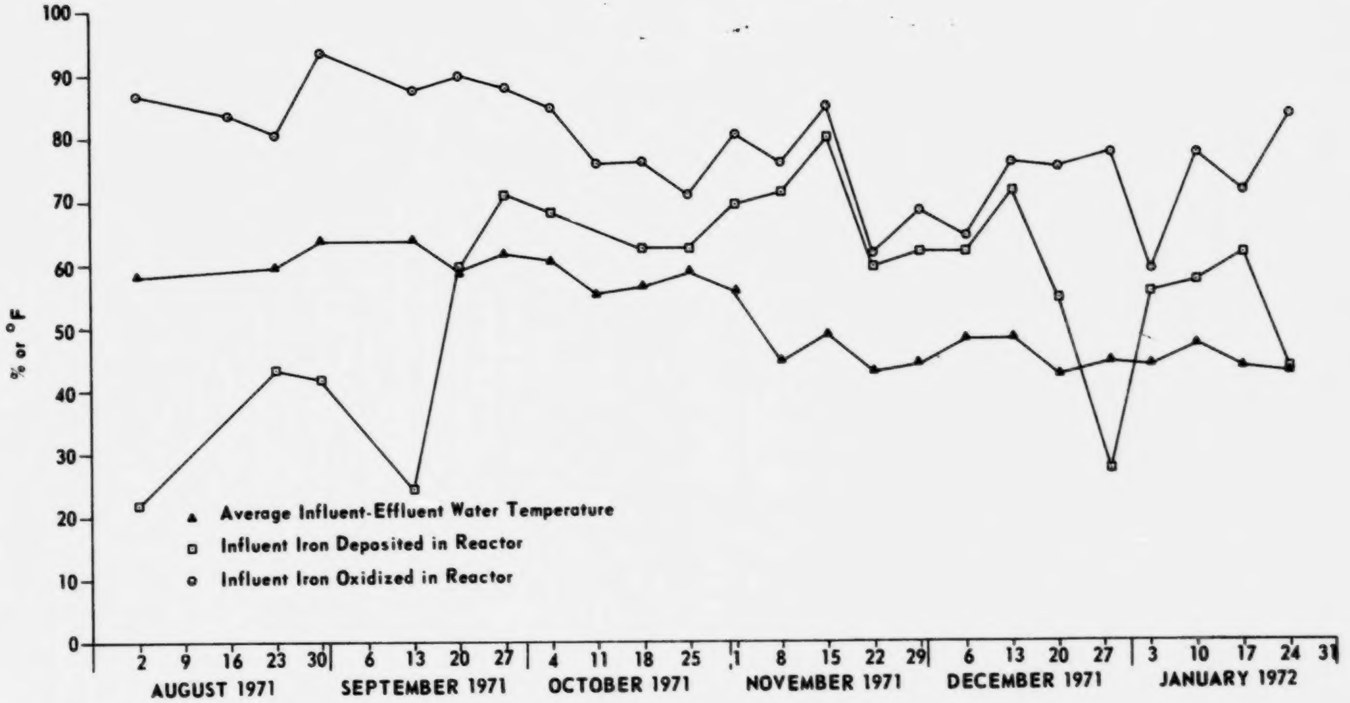
Aeration: Air sparger in stages A and B at approximately 1/10 vol. air/min./volume water.
Surface aeration in stages C-D-E
Agitation: Slow mechanical stirrer by surface aerator.
24 RPM in stages C and D
Water temperature: 61°F
Culture: Strain Z, 93.5 hours old
Raw Water Analyses: Iron (II): 340-358 mg./l.
Iron (Total): 410-420 mg./l.
Sulfate: 1,950 mg./l.
pH: 3.0 - 3.5
Acidity: 1,350-1,500 mg./l.

deviate extensively with flow changes until a flooding condition would develop which probably would not permit sufficient oxidation time. This retention value was attained by batch introduction of a concentrated tracer (sulfuric acid and sodium sulfate were both helpfully used) to the surface of the media and the conductivity of the effluent measured as a function of time. The first increase in conductivity was observed after one minute. The response of the tracer had passed in about 12 minutes.

The overall oxidation performance of the reactor is shown in Figure 23, indicating the extent of iron II reacted varied between 60 and 94 percent.

In comparing deep vat and surface reactor biochemical iron II oxidation, three pragmatic conclusions are made which favor the surface-type reactor:

Figure 23. Surface biochemical reactor performance



1. The surface reactor is least sensitive to disruptive influences, i.e., change of iron II concentration, metabolic inhibitants, flow rates, etc.

2. The surface reactor permits much higher reaction rates than the deep vat systems under continuous flow operation conditions: at least 6000 mg/l/hr versus 50 mg/l/hr.

3. The surface reactor has no power requirements to control water flow, introduce oxygen and carbon dioxide.

The control and removal of the deposit of ferric oxyhydroxides in the surface reactor must be further studied. In the year's operation of the surface unit at the Hollywood Facility, hydraulic flow rates have not been significantly reduced. Water temperature response during winter operation needs further evaluation. Present experience extended into sub-freezing weather permitting feed water to freeze and prevent flow. In studies now in progress, a simple translucent, plastic film cover over the reactor appears satisfactory to permit continued operation in winter weather thus avoiding the cost of introducing any heat to the water.

CALCULATION OF OBSERVED IRON OXIDATION RATES IN PLANT OPERATION

The estimation of the biochemical oxidation rate for ferrous iron in a continuous vat system involves two parameters: (1) iron II concentrations in the influent and effluent, and (2) the available retention time. The latter is controlled by reactor size and flow rate. Other parameters, as pH, nutrients, temperature, agitation, etc., are assumed to be constant and to be maintained at optimum levels. Starting with a vat filled with inoculated raw mine water as a batch operation, the iron II content will decrease as the cell population increases. When continuous flow is initiated, should no further iron oxidation occur, the iron concentration will rise according to the relative concentrations as it approaches the influent concentration. The rate of approach will vary with flow rate and reactor volume. However, with oxidation, a new equilibirium iron II concentration will be established, based on the oxidation rate. A computer program to establish this rate was developed and is listed in Appendix D along with an illustrative calculation. The program provides a time-step integration, which begins with a constant reactant volume having a given initial iron II concentration. Utilizing a specified time interval (as 1, 2 or 5 minutes), the iron input is designated (flow and iron II concentration) and assumes instantaneous mixing. The resulting concentration of iron II will change with the existing oxidation rate, accompanied by the effluent loss corresponding to the stated flow rate. Based on a defined oxidation period such as 30 or 60 minutes, effluent iron II concentration, and an initial oxidation rate of zero, the program will read out the iron content per time interval (in pounds and mg/l), the change in iron content per interval and the cumulative change during this defined period. The program subsequently calculates a first approximation oxidation rate to maintain the defined initial reactor concentration. In a second loop of estimations, the computer utilizes the approximation oxidation rate and, printing out the above interval data, establishes the adequacy of the rate for the defined oxidation period (one to two hours at five minute intervals is usually an adequate simulation period). Should the initial rate estimate not conform within one percent of the equilibirium concentration, a new oxidation rate is estimated and the calculation repeated.

This program, and minor variations thereof, serves as a simulation model for design purposes. It also provides an insight into the capabilities of multi-stage systems. Two important factors must be considered in such an evaluation: (1) the specified or acceptable iron II level of the effluent of the oxidation system, and (2) the iron II level of the raw water influent. In general, the former concentration becomes the equilibrium concentration of the system and, since the oxidation rate is distinctly a function of iron II concentrations, the oxidation rate can become a severely limiting factor for the system. Although this model was designed for the biochemical vat process, it is applicable to the chemical-air oxidation procedure.

The limitations of oxidation rate defined by the equilibrium iron II concentration appears to account for the limitations of the vat oxidation approach and explains the superior results of the surface reactor system. In the latter case, there is no dilution and the oxidation proceeds at the highest prevailing iron concentration as the water contacts the media surface.

The implications of biochemical iron oxidation were included in Section XII, describing the limestone treatment system.

SECTION XIV

SLUDGE CONTROL IN COAL MINE DRAINAGE TREATMENT

INTRODUCTION

Any treatment removing the undesirable ions from mineralized waters creates a by-product which is troublesome and costly to control. In neutralization processes applied to coal mine drainage, a sludge is formed from the separated water-insoluble compounds. Lovell in 1970[64] reviewed the sludge control philosophy basic in these studies. This detail should provide a positive base for developing an optimum sludge control system.

Lovell noted the need in sludge control planning to coordinate the nature and volume of the water to be treated with the site, treatment process, and treatment period anticipated. The solids separation approach must offer control to prevent accidental discharge of inadequate quality waters. A final polishing pond enhances this capability.

Sludge control may incorporate three stages: solid-fluid separation of the sludge, sludge dewatering, and sludge disposal. Two techniques may be satisfactorily employed in the solid-fluid separation: settling lagoons and mechanical clarifier-thickeners. Lagoons are most commonly employed. Although the choice is usually determined by land availability and process economics, there are other more subtle differences which should not be ignored. The process economics do not consistently favor either practice; the guiding factors which define costs will be indicated here.

Solid-fluid separation in a settling lagoon may provide limited sludge storage or permit permanent disposal. Limited storage requires sludge transfer but uses less land. Settling and disposal in the same lagoon necessitates large land areas, but involves minimal design and construction. The method can be quite satisfactory. This approach is the least acceptable aesthetically. Other detailed design, construction, and operation considerations are also necessary.

In contrast, mechanical clarifier-thickener systems require the least land area and provide the least sludge storage. They are more versatile in process control by permitting continuous, automatic sludge removal. Their mechanical nature does not necessarily imply higher costs.

Further dewatering of the settled sludge prior to disposal may be indicated by its voluminous nature, low percentage of solids, and tendency to gel. It may involve some form of in situ evaporative drying or direct water removal by a drying basin, by a filter, or by a centrifuge. The settled sludge usually may be pumped through pipes to the disposal area or to a tanker truck for haulage to the disposal site requiring multiple handling. In either case, handling large volumes of water increases costs. However, after dewatering the solids may be moved by dump truck. Ultimate disposal is usually made into abandoned mines (deep and surface) or to land fill. No broad, feasible use for the impurity solids has been developed.

The settled sludge is difficult to handle and dewater. When collected by pump suction, the gelled or flocculated solid particles do not readily disperse. They move at slower rates than the supernatant water which leads to "rat holing". This phenomenon makes removal of settled sludge from a lagoon bottom difficult and costly. Some sludges tend to gel, while others achieve a plastic clay-like character which resists any flow.

The important physical properties of the sludge are: settling rate, settleability (an arbitrary concept combining settling rate and final settled volume which also incorporates compression phenomena), bulk density, and flowability. All relate to the hydrated character of the sludge. The sludges settle to a solids content between 0.5 and 14 percent by weight. Most are in the lower range. When dried or frozen, the highly hydrated character of the sludge is changed. These hydration properties are not restored upon rewetting. The physical properties of sludges are governed by chemical composition and precipitation processes which tend to be reasonably uniform for a given drainage water and treatment process. However the properties vary widely with the neutralization reagent employed. Hoak[65], Sanmarful and Kountz[66], Duehl[67], and Glover[42] have cited some detail of sludge property variation with different neutralization reagents. Limestone shows the greatest deviation from the other reagents, producing a rapidly settling sludge.

The volume of the settled sludge increases with iron-aluminum concentration of the raw water extending to more than 30 volume percent. The sludge volume level obviously relates to disposal problems and costs; there are advantages in achieving higher density sludges. In addition to control by the initial precipitation reactions, a greater sludge density may be achieved by recycling a portion of the settled sludge to the initial reaction chamber. This approach is utilized in municipal water and sewage treatment and is cited for application to coal mine drainage treatment by Gustafson[13], Lovell[64], Birch[50] and Kostenbader, et al.[15]. A very dense, rapid settling magnetic sludge has been described [Streeter and Lovell[68], and Stauffer and Lovell[69]] when the iron is converted to a magnetite structure. Although of extensive potential, this procedure has not be developed in a feasible system.

The factors which relate density, flowability, and surface character of sludges to their composition and formation, have not been adequately elucidated. Some of these aspects have been considered by Langmuir and Whittemore[70, 71] from a geochemical viewpoint. Obviously the crystal composition, size, surface area, and the nature of its surface charges all affect the sludge behavior. Sludges formed in the pH range of 6.4 to 7.2 appear to settle most rapidly. When feasible, this pH range was sought throughout these studies. Following this principle also permits use of minimum reagent levels. These researchers find the sludges to be primarily iron oxyhydroxides. The natural "yellowboy" is poorly crystallized goethite (orthorhombic $Fe_2O_2 \cdot H_2O$) and amorphorous ferric oxyhydroxides. Sludges prepared by synthetic neutralization yield a mixture of poorly crystallized goethite and lepidocrocite. It is probable that aluminum and other ions are detrimental to crystallization. In pure chemical systems, superior crystallization results. When using sodium bicarbonate to neutralize these solutions, it appears lepidocrite initially forms then reverts to goethite. When seeding is utilized, a reversal of forms occurs with goethite appearing first but a mixture still results. Sludges resulting from hydrolysis after biochemical iron oxidation of drainage waters are poorly crystallized, amorphorous, ferric oxyhydroxides.

SLUDGE SETTLING RATES

Three types of settling rate phenomena have been observed for the CMD slurries developed at Hollywood.

Type I - The sludge settles rapidly and uniformly with a distinct interface and a clear supernatant liquor. This response results from reaction with strong alkalies (hydrated and pebble lime, sodium hydroxide, and sodium carbonate) when the mine water has a typical or high iron and aluminum content. The settling rate curve has a classical shape, i.e. the sludge volume decreases sharply during the first 10-30 minutes, asymptotically approaching a relatively constant terminal volume within 24 hours. See Figures 24 and 25.

Type II - This behavior occurs when very low iron- and aluminum-containing waters are treated with strong alkalies. Tyler Run waters containing 1 ppm iron and 30 ppm aluminum are illustrative. The amount of precipitate formed is quite small and resists flocculation. The flocs tend to remain diffused and form a nearly stable suspension. Complete settling may not occur for several days. Even then the settled sludge might not exceed 0.1 weight percent solids. The water may retain a slight turbidity. Rates are difficult to evaluate.

Figure 24. Sludge settling rate - solid-fluid separation influent proctor no. 1 water

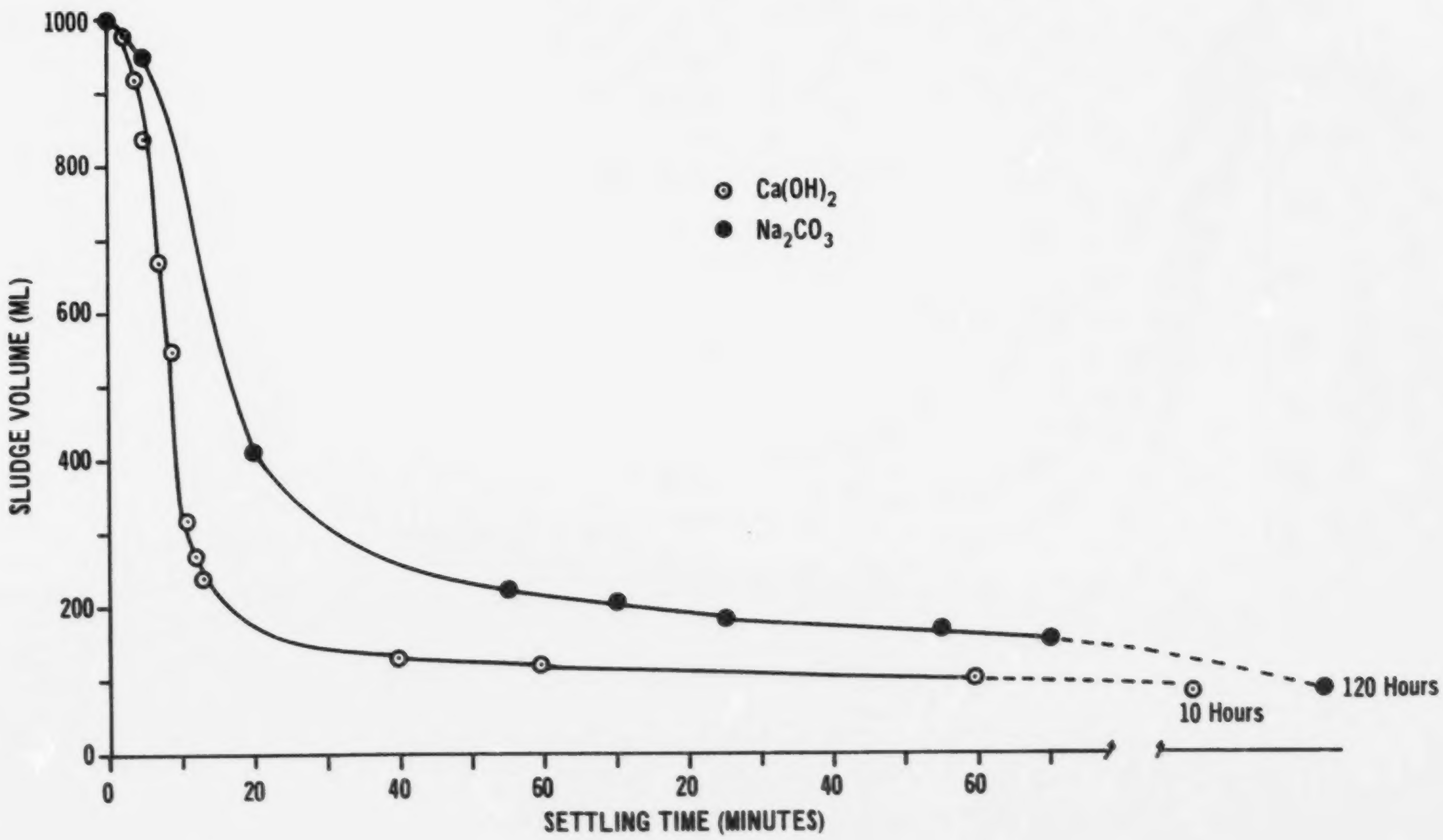
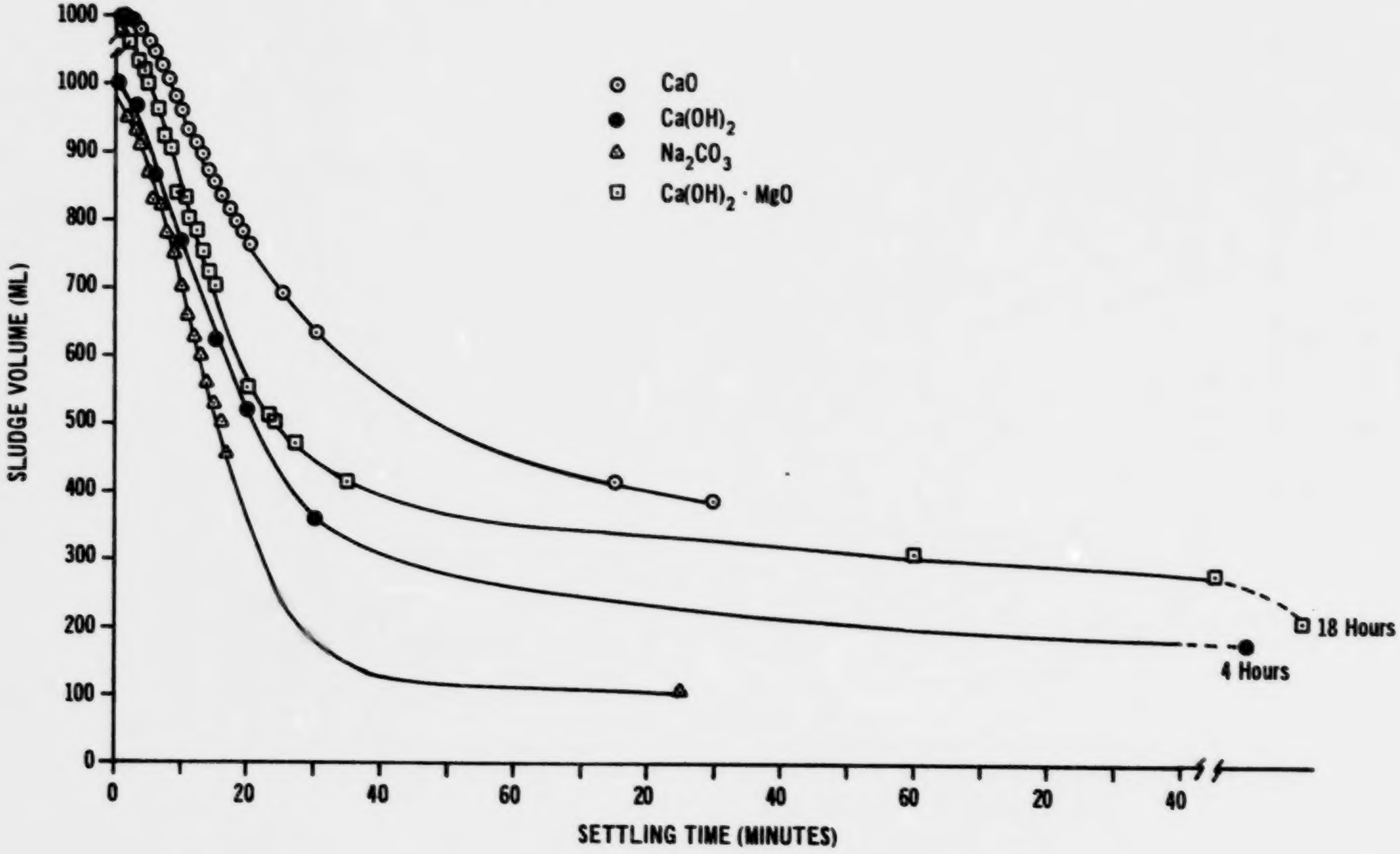


Figure 25. Sludge settling rate - solid-fluid separation influent proctor no. 2 water



Type III - A two-phase solids system is developed. The major portion (to 90 percent) of the solids settle very rapidly and can best be measured as they collect on the bottom of the test device. The particles are not gelatinous flocs and do not appear to "grow" significantly in size. This build up is an asymptotic reciprocal of Type I behavior. The bottom collection usually approaches its maximum level within 10 minutes. This volume is used as the sludge volume. This build-up rate can be measured but with less certainty. The values reported are conservative. The supernatant liquid is turbid. This settling behavior appears to be independent of pollution loading of the feed drainage. Sometimes an interface will form slowly to yield a clear supernatant. This behavior is similar to Type II slurries, requiring about 24 hours to settle. In other cases the solution will decrease in turbidity but remain cloudy. This behavior results with limestone neutralization. The residual turbidity is attributed to the presence of near-colloidal solids. The behavior has also been observed with sodium carbonate and sodium hydroxide when continuing reactions (both neutralization and iron oxidation) are experienced. When observed with these reagents, the behavior is a symptom of improper reaction control.

Solids separation from all three types of settling behavior can be improved by the use of flocculants, by sludge recycling to increase the total solids content of the suspension, or by "filtering" the floc through a deep sludge blanket such as can be maintained in a thickener. The sludge blanket in a thickener can be improved by weighting it with loading agents such as pulverized limestone, fine coal and silica.

The standard settling rate test utilized was performed by filling a 1,000 ml. graduate with a freshly prepared slurry. After gentle mixing (not stirring) to insure uniform distribution of the particles and to permit floc growth to be initiated, the interface volume between the sludge and clear supernate was recorded as a function of time. The settling rate was determined from the free settling portion of the data (i.e., prior to the start of sludge compression).

A summary of sludge settling data from various plant tests are presented in Table 24. No data resulted from the sludges exhibiting Type II behavior. The observed settling rates ranged from 1.2 to 5.4 ft/hr. The higher values resulted with limestone treatment but yielded a turbid supernatant liquor. Plant data did not indicate rate trends among the other reagents. The sludge settling rate seemed to approach a maximum with increasing iron-aluminum levels

Table 24. SUMMARY OF SLUDGE SETTLING DATA FROM HOLLYWOOD PLANT TESTS

Water Source	Total Acidity mg/l	Fe^{++} mg/l	Al mg/l	Final pH	Alkali Used	Sludge Settling Rate ft/hr	Suspended Solids in Neutralized Water mg/l (aver.)	Sludge Settled Volume % of original	Solids in Settled Sludge wt. %	Calculated Solids in Settled Sludge (T.A. + $2Fe^{++}$)
Tyler Run	114	4	15	6.26	NaOH	--	50	2.2	0.23	--
Bennett's Branch	120	29	15	6.51	NaOH	--	91	--	--	--
Bennett's Branch	206	35	17	6.95	CaO	--	96	4.5 + T	0.21	0.61
Bennett's Branch	206	35	17	6.95	CaO	--	96	2.5 + T	0.38	1.10
Proctor No. 1	502	80	80	8.30	CaO	3.49	350	10.0	0.35	0.66
Proctor No. 2	1,707	420	210	7.46	CaO	1.02	1,365	29.6	0.46	--
Tyler Run	78	1	15	7.00	Soda Ash	--	--	--	--	--
Bennett's Branch	86	10	5	8.03	Soda Ash	--	37	1.0	0.37	1.06
Proctor No. 1	397	50	50	7.38	Soda Ash	2.96	290	5.0	0.58	1.00
Bennett's Branch	79	18	10	8.00	Soda Ash	--	55	0 + T	--	--
Proctor No. 1	328	57	57	7.60	Soda Ash	3.95	252	3.0	0.84	1.47
Proctor No. 1	302	52	75	7.78	Soda Ash	1.87	178	3.0	0.59	1.35
Proctor No. 2	1,398	241	205	7.31	Soda Ash	1.78	2,318	11.0	2.11	1.71
Proctor No. 2	1,174	118	59	7.38	Soda Ash	3.00	--	8.0	--	1.76
Proctor No. 1	405	69	62	7.25	$Ca(OH)_2$	4.72	848	8.0	0.06	0.68
Proctor No. 1	1,380	317	191	6.75	Limestone	5.40+T	1,692	3.0	5.63	6.69
Tyler Run	104	<1	12	7.75	Pul.Limestone	4.68+T	104	0.2	5.2	5.2
Proctor No. 2	1,401	322	191	7.75	Pul.Limestone	4.10+T	1,356	2.0	6.78	10.3
Proctor No. 2	1,305	250	191	7.1	$Ca(OH)_2$	1.47	1,370	18.0	0.76	1.0
Proctor No. 2	1,360	266	191	8.2	Dol.Hydrate	1.95	1,990	18.0	1.11	1.05

NOTE:

"+T" - Indicates supernatant liquor was turbid.

in the 100-200 ppm range then decreased. At low metal levels floc growth was slow while at levels above 200 ppm the growing flocs approached compression and settled more slowly. There was no distinct trend measured during plant tests in settling rates with pH (6.3-8.3) variation, if the neutralization-oxidation reactions were complete. However, laboratory tests indicate maximum rates near pH 6.8. With prior reports of this dependence, the relationship between settling rates and pH precipitation levels might be considered further. Continuing reactions literally "muddied" the water. The density of the settled sludge does not directly reflect either metals content of the raw water or settling rates of the sludge.

USE OF POLYELECTROLYTES AS SETTLING AIDS

Polyelectrolytes (usually polyacrylamide polymers of various molecular weights and structures) have found universal application in enhancing the settling rates of all types of slurries, including most coal mine water sludges. They enhance floc growth and decrease turbidity. They are available in a range of ionic charges, molecular weights, and other characteristics which govern their performance. There is no evidence that these polymers increase final sludge densities either in bench or plant tests. This was evident in sludge from Tyler Run waters which have low suspended solids containing predominantly aluminum hydroxide. At very high polymer dosage, a "balling" phenomenon occurs with these sludges which may have the practical response of increasing density. The polymers do not decrease residual soluble iron levels in treated waters as has been suggested[72]. Since their usage does aid uniform plant operation and reduce the size of equipment, an evaluation of several types of polyelectrolytes was made. Previous pilot studies[75] involving several preparations found non ionic Superfloc 84 and Nalcolyte 67 effective at dosages of about 10 ppm.

Representatives of several polyelectrolyte producers carried out evaluation of their flocculants on fresh sludge at the pilot plant. These individuals, being knowledgeable regarding specific formulations, were most qualified to make preliminary screening tests. Although the use of these formulations is simple, the techniques must be carefully established. Considered were: selection of the most effective reagent, preparation of the reagent solution, and reaction parameters as dosage, dilution and mixing procedures.

These polymers consist of long length molecules. Their response is governed by the reaction of the polymer and the particle surface, thus contacting methods between the polymer and the suspended particles are critical. Excessive agitation (as passing a solution through a high speed mixer or pump impeller) can shear polymer molecules and destroy their effectiveness.

The data presented are not intended to compare products nor imply that other products may be unsatisfactory. High molecular weight polymers can be effective flocculants at dosage levels of 0.5 to 2.0 ppm of neutralized mine water. Those polymers designated non ionic or weakly anionic were most responsive. The settling rates with the polymers ranged from 4 to 16 feet per hour. The optimum dosage level increased with the suspended solids level. In recycle sludge systems, flocculant dosage levels should be based on the ppm polymer/lb of dry solids. It cannot be assumed that polymers associated with settled recycle particles remain effective. Excessive agitation can destroy polymer-formed flocs. The best flocculant and dosage for any particular sludge must be established experimentally, although the data presented here should limit the evaluation. No obvious sensitivity of the polymers to pH was indicated over the range 6.0-9.0 normally experienced.

In testing American Cyanamid reagents, the diluted flocculant was added to 1,000 ml slurry in a graduated cylinder during the first three of fifteen slow, mixing strokes using a perforated disk plunger. Timing was initiated at the completion of mixing. The sludge interface location was estimated for Bennett's Branch and Tyler Run water due to its indefinite nature. Supernatant turbidity was established with a Helige Tubidimeter.

The evaluations with different waters and neutralization reagents are detailed in Tables 25 through 31. Those recommended by their respective companies were: Superfloc 127, Decolyte 930 and Atlasep 1A1.

Some polymers were utilized in plant tests involving the thickener, Densator, settling lagoon, filter and centrifuge operations. Their response confirmed bench scale testing and are described in the appropriate sections of this report. Somewhat higher dosages than those indicated by bench tests were desirable during the plant tests. The ability to increase the capacity several fold of a solid-fluid separation unit by using polymers was demonstrated.

PRIMARY SOLID-FLUID SEPARATION OF TREATED COAL MINE DRAINAGE

Thickeners

Gravity settling of suspended solid particles is divided into clarification and thickening[73]. The former term is applied to dilute pulps as illustrated by Type II mine drainage sludges. The devices utilized are characterized as to means for removal of the settled solids. Unit sizing is the primary consideration. The size is established empirically, based on influent volume and particle levels and characteristics. Particle settling rates establish the surface area required

Table 25. ANALYSES OF MINE DRAINAGE WATERS EVALUATED WITH SUPPERFLOC* REAGENTS

Water designation	Initial pH	Acidity (mg/1 $CaCO_3$)	Iron (II) mg/1	Aluminum mg/1	Slurry pH[a]
Tyler Run	3.6	65	5	20	6.6
Bennett's Branch	3.7	230	38	30	6.7
Proctor No. 1	3.4	463	64	74	6.6
Proctor No. 2	2.9	1765	320	213	6.7

[a] Following neutralization with $Ca(OH)_2$. Flocculation studies were made at this pH level.

Reagents* Evaluated[a]

SF-16	SUPERFLOC 16 Flocculant
SF-127	SUPERFLOC 127 Flocculant
AF-550	AEROFLOC 550 Reagent
SF-202	SUPERFLOC 202 Flocculant
SF-210	SUPERFLOC 210 Flocculant
SF-212	SUPERFLOC 212 Flocculant
SF-214	SUPERFLOC 214 Flocculant
S-3645	Reagent S-3645 Flocculant
SF-310	SUPERFLOC 310 Flocculant
S-3686	Reagent S-3686 Flocculant
S-3698	Reagent S-3698 Flocculant
S-3700	Reagent S-3700 Flocculant
GPG	AEROSOL GPG Surface Active Agent
S-3665	Reagent S-3665 Surfactant
S-3699	Reagent S-3699 Flocculant

*American Cyanamid Company, Inc.

[a] Utilized as 0.1% by wt. aqueous solutions. The required volume of this solution was diluted to 30 ml before reacting with the test slurry.

Table 26. SLUDGE SETTLING RATES WITH SUPERFLOC POLYELECTROLYTES. WATER ORIGIN - BENNETT'S BRANCH

Flocculant	Dosage (ppm)	Floc Size	Settling Rate	Flocs Suspended after 15 minutes
Control	-	V. Small	V. Slow	75%
SF 127	0.5	Large	Fast	0%
	0.25	Large	Fast	0%
	0.1	Medium Large	Medium	20%
AF 550	1.0	Small	Slow	40%
	0.5	Small	Slow	40%
SF 202	0.5	Large	Fast	5%
	0.25	Large	Fast	15%
	0.1	Medium	Slow	60%
SF 210	0.5	Large	Fast	5%
	0.25	Large	Fast	15%
	0.1	Medium	Slow	50%
SF 214	0.5	Large	Fast	5%
	0.25	Large	Fast	15%
	0.1	Medium	Slow	50%
SF 310	2.0	V. Small	V. Slow	60%
	3.0	Small	Slow	40%
S-3686	2.0	V. Small	V. Slow	75%
	3.0	Small	Slow	50%
S-3698	2.0	V. Small	V. Slow	75%
S-3699	2.0	V. Small	V. Slow	75%
SF 16	0.25	Medium	Medium	40%
SF 212	0.25	Large	Fast	15%
	0.1	Medium	Slow	50%
S-3645	0.25	Large	Fast	5%
	0.1	Medium	Slow	50 %

[a] American Cyanamid Company

Table 27. SLUDGE SETTLING RATES WITH SUPERFLOC POLYELECTROLYTES. WATER ORIGIN - PROCTOR NO. 1

Flocculant	Dosage (ppm)	Time to 200 ml Sludge Volume (seconds)	Pulp Volume after 33 min (ml.)	Turbidity of Effluent after 33 min or 15 min* (ppm)	Calc. Settling Rate (ft/hr)
Control	-	772	105	40	4.67
	-	670	100	225*	5.37
SF 16	0.15	503	98	22	7.14
	0.05	580	102	125*	6.20
SF 127	0.15	220	70	22	16.14
	0.05	410	89	40*	8.76
	0.05	342	90	50*	10.50
AF 550	0.3	903	115	22	3.98
	0.10	606	110	105*	5.33
SF 202	0.15	428	100	22	8.40
	0.05	525	105	70*	6.18
SF 210	0.15	268	90	22	13.4
	0.05	420	98	50*	7.80
SF 214	0.15	398	100	22	9.00
	0.05	458	100	70*	7.08
S-3645	0.15	340	88	22	10.60
	0.05	475	90	90*	6.84
SF 310	0.6	545	102	30	6.62
	0.20	677	100	240 *	4.79
S-3699	0.6	597	102	30	6.00
	0.20	611	100	205*	5.30
SF 212	0.05	444	99	50*	7.26
S-3686	0.20	600	100	125*	5.98
S-3698	0.20	702	100	185*	5.14
S-3700	0.20	699	100	165*	5.15

[a]American Cyanamid Company

Table 28. SLUDGE SETTLING RATES WITH SUPERFLOC POLYELECTROLYTES. WATER ORIGIN - PROCTOR NO. 2

Flocculant	Dosage (ppm)	Time to Sludge Volume, seconds					Pulp Volume after 40 min (ml)	Overflow Turbidity after 40 min (ppm)	Calc. Settling Rate (from 500 ml-ft/hr.)
		800 ml	600 ml	500 ml	400 ml	300 ml			
Control	-	514	963	1313	1810		340	50	1.71
SF 16	0.5	283	650	1044	1456		300	30	2.16
SF 127	0.5	126	322	515	818	1372	235	30	4.37
	0.25	315	710	997	1478		290	30	2.26
AF 550	1.0	350	892	1231	1863		310	30	1.87
	2.0	504	898	1267	2023		365	30	1.78
SF 202	0.5	385	824	1183	1752		350	30	1.93
SF 210	0.5	217	516	880	1221	2192	295	30	2.06
	0.25	354	801	1167	1722		330	30	1.93
SF 214	0.5	324	734	1078	1576		335	30	2.09
	0.25	387	857	1240	1784		340	30	1.82
S-3645	0.5	232	569	851	1291	2090	290	30	2.65
	0.25	343	765	1106	1614		320	30	2.04
SF 310	4.0	350	641	874	1341		320	50	2.57
	2.0	450	807	1097	1581		345	50	2.06
S-3699	2.0	494	899	1220	1737		350	50	1.85
SF 212	0.25	361	820	1212	1761		335	30	1.86
S-3686	4.0	396	697	943	1381		320	50	2.39
S-3698	4.0	521	935	1254	1768		345	50	1.80
S-3700	4.0	338	620	845	1303		320	50	2.65

[a] American Cyanamid Company

Table 29. SLUDGE SETTLING RATES WITH DECOLYTE POLYELECTROLYTES

Water Source: Tyler Run - neutralized to about pH 9.0 with hydrate lime slurry

Typical Analysis:
pH - 3.42
Acidity - 126.7

Iron (II) - <1 mg/l
Iron (T) - 3.4 mg/l

Aluminum - 19 mg/l
SO_4 - 259 mg/l

Evaluation of Flocculant Type

	Blank	DECOLYTE 710	DECOLYTE 720	DECOLYTE 930	DECOLYTE 940	DECOLYTE 950
Dosage, ppm	0	1	1	1	1	1
Polymer Ionic Character		+	+	-	-	-
Floc Size	Small	Small	Small	Large	Large	Medium
Settling Rate[1]	Slow	Slow	Slow	Fast	Fast	Medium
Relative Performance Rating (1 = Best)	6	5	4	1	2	3

Evaluation of Flocculant Dosage (DECOLYTE 930)

	0 ppm	0.25 ppm	0.50 ppm	0.75 ppm	1.5 ppm	2.0 ppm
Floc Size	Small	Large	Large	Large	Large	Large
Floc Settling[1]	Slow	Fast	Fast	Fast	Fast	Fast
Relative Settling Performance (1 = Best)	6	5	4	3	2	1

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[1] The level and nature of floc produced from this water does not permit interface evaluation.

Table 29 (continued). SLUDGE SETTLING RATES WITH DECOLYTE POLYELECTROLYTES

Water Source: Proctor No. 2 - neutralized with hydrated lime slurry to pH 8.3
Typical Analysis:
pH - 2.81
Acidity, ppm - 1689
Iron (II), ppm - 440
Iron (T), ppm - 498
Aluminum, ppm - 182
SO_4, ppm - 2510
Flocculant Dosage: 0.75 ppm

	Sludge Interface Level (ml. 0.0017 ft/ml)					
Time (mins.)	Blank	DECOLYTE 710	DECOLYTE 720	DECOLYTE 930	DECOLYTE 940	DECOLYTE 950
0	1000	1000	1000	1000	1000	1000
5	970	960	970	950	950	970
8	925	930	930	900	930	940
10	880	880	880	840	890	910

Slurry Origin: Oxidation Tank effluent using Proctor No. 2 water treated with calcined lime
Water Analysis: pH - >9.0
Iron (T) - <1.0 mg/l

	Sludge Interface Level DECOLYTE[a] 930 Dosage, ppm								
Time (mins.)	Blank	0.25	0.50	0.75	1.0	1.25	1.50	1.75	2.0
0	500 ml	500 ml	500 ml	500 ml	500 ml	500 ml	500 ml	500 ml	500 ml
1	480	440	350	400	390	320	300	270	250
2	455	330	260	280	280	230	220	205	180
3	420	260	210	230	230	200	190	180	160
4	370	220	185	190	190	170	160	155	150
5	320	200	170	170	170	160	155	150	145

[a] Diamond Shamrock Corporation

Table 30. SCREENING TESTS WITH ATLASEP FLOCCULANTS

Polymer Type	Control	1N Nonionic	1A1 Weakly Anionic	2A2 Moderately Anionic	3A3 Strongly Anionic	105C Cationic
	Relative Response[c]					
Tyler Run - pH 8.7	3	2	1			
Proctor No. 1 - pH 6.9	3	1	1			
Proctor No. 1 - pH 7.9[d]	4	1	1	2	3	
Proctor No. 2 - pH 7.5	5	1	1	2	4	3
Proctor No. 2 - pH 7.8[e]	5	1	1	2	3	4
Proctor No. 2 - pH 8.3[e]	4	2	1			

Table 31

Sludge Settling Rates with ATLASEP IN* Flocculant

Water Source: Proctor No. 2 neutralized with hydrated lime slurry

Dosage, ppm	Settling Rate, ft/hr.
0	0.85
0.25	2.25
0.50	4.10
1.0	8.04
1.5	11.7

[a] Atlas Chemical Industries, Inc. , ultra-high molecular weight polyelectrolytes.

[b] Jar tests in which flocculant was added as 0.01% solution and mixed for 15 seconds. 600 ml. in 1,000 ml. beaker. Dosage - 1 ppm.

[c] Based on floc formation characteristics, sludge settling rate and supernatant clarity.

[d] Neutralized with limestone, all others with hydrated lime slurry.

[e] Densator recycle sludge, diluted to about 10,000 mg/l suspended solids. Ferrous sludge.

to provide the desired water rise velocity (ft/hr or gal/min/ft^2). The clarified effluent rise velocity must be less than the settling rate of the slowest moving suspended particle. A rise velocity, which is much lower than the particle settling rate, may be required with some slurries to provide adequate retention time for floc growth.

The depth of the thickener must comprise four zones: the clear supernatant, the feed region, and the transition and the compression zones. The clarified zone contributes a safety factor against discharge of turbid waters (usually at least two feet) while the compression zone aids in producing a more dense sludge. Observations during these studies indicate the advisability of introducing the feed slurry from some mine waters into the transition zone to gain a filtration effect of the settling sludge blanket. At the bottom of the compression zone, a slow moving rake moves the sludge to a removal port and prevents non-uniform, excessively dense sludge build up. Typical unit areas for many suspensions range between 50 and 150 (ft^2/ton/day) for influent levels containing one or two percent suspended solids[73]. These values are significantly less than found by Dorr Oliver (109-4000)[75] and in these studies for mine drainage sludges.

Hollywood Thickener--

The device utilized is atypical in its design as a thickener since in addition to this application, studies of the sludge recycle concept were to be made[13] with this unit. The studies with this latter mode of operation are described in Section VIII. The unit is ruggedly constructed and was not prone to mechanical problems. It performed both types of functions satisfactorily; provided means for good control, sampling and observation capability. The calculated flow rate parameters are given in Table 32, while related sludge parameters are given in Table 33.

In this thickener, the primary influent zone serves as a feed well, regulating flow surges and providing additional reaction time. The secondary zone permits flocculant addition and provides slow agitation for floc formation. This design permits the introduction of the feed slurry below the sludge blanket into the transition zone, if desired.

These atypical aspects of this thickener's design reduces the overall capacity. In the solids separation zone (feed solids and transition regions) the downflow settling of the floc proceeds along with the upflow rise velocity of the clarified water. The feed well extends ten and one-half feet below the overflow level. This provision permits an upflow sludge filtration region of eight feet when the sludge blanket is maintained two feet below the overflow. When the

Table 32. THICKENER FLOW RATE PARAMETERS

Zone	Zone capacity gallons	Retention time minutes			Flow velocity feet per hour			Gallons/minute/ft^2		
		100 gpm	200 gpm	350 gpm	100 gpm	200 gpm	350 gpm	100 gpm	200 gpm	350 gpm
Primary (Downflow)	3,730	37.3	18.6	10.7	8.46	16.9	29.4	1.05	2.10	3.68
Secondary (Downflow)	3,730	37.3	18.6	10.7	8.46	16.9	29.4	1.05	2.10	3.68
Compression Zone	13,550 1,130[a]	See Table 33 for Compression Zone Data								
Solids Separation (Upflow)	29,200	292.0	146.0	83.2	2.06	4.12	7.2	0.253	0.506	0.885
Total	51,340	513.0	257.0	147.0						

[a] Sludge Collection Zone

Table 33. SLUDGE RATE PARAMETERS FOR 25-FOOT THICKENER

	Influent Solids Level mg/l								
	500			1000			2000		
	Flow Rate - gpm								
	100	200	350	100	200	350	100	200	350
Solids in Underflow %	Retention Time-Compression Zone-hours[a]								
2	98	49	28	49	25	14	25	12	7
4	196	98	56	98	49	28	49	25	14
6	295	147	84	147	73	42	73	37	21
8	392	196	112	196	98	56	98	49	28
	Sludge Generation (Equilibrium Blow Down) Rate-gpm								
2	2.5	5.0	8.75	5.0	10.0	17.5	10.0	20.0	35.0
4	1.25	2.5	4.38	2.5	5.0	8.7	5.0	10.0	17.5
6	0.83	1.66	2.91	1.67	3.33	5.8	3.3	6.7	11.7
8	0.63	1.25	2.19	1.25	2.50	4.4	2.5	5.0	8.8
	Dry Solids Production Rate								
lbs/min.	0.42	0.84	1.46	0.84	1.68	2.92	1.67	3.34	5.85
lbs/day	605	1210	2102	1210	2420	4204	2420	4840	8408
lbs/ft^2/ 24 hrs	1.23	2.47	4.29	2.47	4.94	8.58	4.94	9.88	17.16

[a] Calculated from compression zone volume of 14,700 gallons and bottom area of 490 ft^2. Sludge depth from bottom of secondary zone - 5 feet.

sludge level is carried above the bottom of the secondary zone wall, a desirable upflow filtration of fines through the blanket occurs and can function to remove near-colloidal material from the rising water column. The thickener was satisfactorily operated under these conditions during several limestone neutralization runs resulting in good effluent clarity. This actual performance deviated significantly from bench scale settling tests with the same feed which failed to yield clear effluents, even with some flocculant additions. The use of thickener designs which include deep feed wells thus permitting upflow sludge blanket clarification is deemed to be an important discovery.

The upflow velocities of this thickener ranged from 2.06 ft/hr at 100 gpm to 7.2 ft/hr at 350 gpm. The measured settling rates of freshly precipitated sludges ranged between one and six ft/hr with a maximum of 16 ft/hr with flocculants. More typical sludge settling rates were nearer two ft/hr. This size thickener would function satisfactorily with flows near 100 gpm; however, when flocculants were used the rates could be expanded about three times and still maintain a clear effluent. The functioning of the primary and secondary zones were quite adequate for the maximum flow (350 gpm) to which the unit was subjected. In some tests, with Type II, slow settling sludges, and with the use of pulverized limestone as a. reagent, excessive turbidity in the effluent did occur.

Densator Performance as a Thickener--

The Densator was operated as a thickener. Typical performance data for the various conditions are shown in Tables 34 and 35. All data utilized the following flowsheet. Limestone reactions introduced a pulverized slurry into the flash mixer since the rotary limestone reactor effluent could not be introduced to the thickener.

Raw Water → Flash Mixer → Oxidation Tanks → Thickener
↙ ↘
Alkali Slurry ——↑ Dewater ← Underflow Effluent
↓
Settling Lagoon
of Stream

Good clarification was maintained at flow rates less than 125 gpm when 1 to 2 ppm of flocculant were used. Flocculant additions were essential. Higher flow rates presented difficulty in keeping the sludge blanket level below a 2-foot depth from the overflow. This suggests rising velocities greater than about 3 ft/hr should not be exceeded.

Sludge thickening was erratic with underflow slurries ranging from 0.7 to 7 percent solids. The lower levels resulted from low acidity

Table 34. TYPICAL THICKENER PERFORMANCE DATA

	Water Source										
	Tyler Run		Bennett's Branch			Proctor No. 1			Proctor No. 2		
	Neutralization Reagent										
	CaO·MgO	Pul. $CaCO_3$	$Ca(OH)_2$	CaO·MgO	Pul. $CaCO_3$	$Ca(OH)_2$	CaO·MgO	Pul. $CaCO_3$	$Ca(OH)_2$	CaO·MgO	Pul. $CaCO_3$
INFLUENT											
Flow Rate - gpm	62	100	140	62	90	100	85	100	106	257	120
Suspended Solids - mg/l	64	75	190	160	780	675	750	1500	1860	1600	1870
pH	7.7	7.6	8.6	7.4	7.5	8.3	7.5	7.0	7.1	8.0	8.3
Ferrous Iron - mg/l	1	1	1	1	1	4	2	10	2	9	1
Alkalinity - mg/l	26	32	23	15	15	9	17	10	24	914	84
Temperature - °F	50	49	51	51	52	50	49	45	60	45	51
Settling Rates - ft/hr	0.5	*	1.2	1.5	*	3.5	1.9	*	2.2	1.8	1.9
Volume of Settled Sludge - %[a]	20		4.5	0.5-4.0		12	5		20	10	20
Flocculant Dosage - mg/l (Atlasep 1A1)	1-2	1-2	1-2	1-2	1-2	1-2	1-2	1-2	1-2	NONE	1-2
OVERFLOW EFFLUENT											
Suspended Solids - mg/l	15	30	5	4	4	10	27	18	19	683	51
Soluble Ferrous Iron - mg/l	1	1	1	1	1	1	1	4	1	1	1
Alkalinity - mg/l	20	32	22	10	16	85	5		65	1632	66
pH	7.5	7.5	8.3	8.2	7.5	8.2	8.5	7.1	7.2	7.2	8.4
UNDERFLOW											
% Solids, wt/vol	0.8	1.6	1.96	2.1	2.5	6.7	2.5	4.8	2.7	1.3	7.19

[a] Settled sludge volume is percentage of original volume of influent after two hours settling time.

* An average value of 2.0 ft/sec which will yield a moderately clear effluent. Probably acceptable ecologically since the turbidity would be caused by $CaCO_3$ fines. A conservative value of 20 ft/sec assumed for settling rate of bulk of precipitate.

Table 35. OBSERVED THICKENER PERFORMANCE - SIZING DATA

	Water Source										
	Tyler Run		Bennett's Branch			Proctor No. 1			Proctor No. 2		
	Neutralization Reagent										
	CaO MgO	Pul. $CaCO_3$	$Ca(OH)_2$	CaO MgO	Pul. $CaCO_3$	$Ca(OH)_2$	CaO MgO	Pul. $CaCO_3$	$Ca(OH)_2$	CaO MgO	Pul. $CaCO_3$
Feed Dilution[a]	15625	13333	5263	6250	1282	1481	1333	667	538	625	535
Final Dilution[a]	125.0	62.5-25.0	51.0-36.8	47.6-31.3	40.0	14.9	40.0	20.8	37.0	67.9	13.9
Maximum - Feed Retention Time - hrs.	11.7	8.6	6.2	11.7	9.5	8.6	9.8	8.6	8.1	3.4	7.3
Compression Zone - Sludge Retention Time - hrs.	494	523-1307	180-249	518-789	87	243	96	78	34	8	79
Unit Area[b] $ft^2/T/day$	41,230	8,845-884	5,790	5,512	828-83	558	907	430-43	303	412	366-37
Sludge Production $lbs/ft^2/day$	0.05	0.23	0.34	0.36	2.4	3.6	2.2	4.6	6.6	4.8	5.4

[a] Weight ratio of liquid to solids based upon suspended solids determination

[b] $$\text{Unit Area} = \frac{1.333(\text{Feed Dilution}\frac{\text{wt. liquid}}{\text{wt. solids}} - \text{Final Dilution}\frac{\text{wt. liquid}}{\text{wt. solids}})}{\text{Suspended solids settling rate ft/hr}}$$

waters at high flow rates. A typical underflow solids content was about 2 percent. The more dense sludges developed when pulverized limestone was used as the neutralizing agent. This experience was consistent with the results of bench settling tests.

Other observations from the operation of the Densator as a thickener are indicated in part by the data in Tables 34 and 35:

1. Pulverized limestone, as a neutralizing agent, often resulted in turbid overflows which were difficult to control. Pulverized limestone produced a greater quantity of neutralized solids in the influent than with a similar alkaline equivalent of hydrated lime.

2. Reagents with slower neutralizing reactions rates, as dolomitic hydrate and limestone, continued both neutralizing and oxidizing reactions in the thickener. These continuing reactions were apparent from decreases in ferrous iron concentration and increases in pH in the thickener effluent. A thickener can be expected to provide important secondary reaction time to insure specification effluents, but with possible turbidity problems.

3. A sludge level maintained above the bottom of the feed well provides for particle entrapment during upflow through the sludge blanket. This entrapment improves effluent clarity when difficult dispersions occur, as in limestone neutralization.

4. The thickener bottom rake action produces a dewatering action on the settled sludge. Some underflows attained a solid content of five to seven percent after 24 hours of settling time, whereas the same slurry, settled under quiescent conditions, never exceeded two percent solids after 48 hours settling time.

5. Waters having low levels of pollutant loading, produce low settled solids density and require large unit areas. In plant design, serious consideration should be given to the use of a second stage thickener possibly with flocculant addition, to attain a more dense sludge product.

6. Sludge removal control followed either slow continuous pumping or timed-interval blow-down, utilizing automatic pneumatic valves in the system. The test schedule was too short to establish optimum sludge storage and blow-down considerations.

Recommended Design Parameters for Thickeners--

The thickener studies provided information for establishing the design criteria for effective clairfying and thickening of mine slurries. An optimum design would provide the following features:

1. Clarity of overflow with all types of slurries.
2. An underflow of high density solids suitable for secondary dewatering.
3. Provision for sludge recycling to achieve underflow solids content of at least five percent.
4. Provision for second stage thickening, if necessary.
5. Provision for upflow of suspensions through sludge blanket if colloidal solids are present.
6. Provision for settled sludge surge capacity to permit economical sizing and operation of secondary dewatering devices or disposal.
7. Provision for flocculant addition to improve and control sludge settling rates.

Design parameters required to attain the above conditions were established and are listed in Table 36.

Estimated Thickener Operation Costs--

Capital costs of thickening equipment can vary appreciably. For preliminary estimation purposes, the installed equipment costs for various sizes of thickeners are presented in Table 37. The thickener sizes selected are based upon a minimum sludge settling rate of 1.5 ft/hr and a rising velocity of about 0.2 gpm/ft^2. The settling rate largely controls the size of the thickener required. The effect of sludge settling rate on equipment costs is shown in Figure 26. The natural settling rate of most slurries can be increased several fold with polyelectrolyte flocculants. Figure 26 permits estimation of the potential of flocculant usage in determining thickener sizing and costs.

Estimates of thickening costs for three sizes of plants are shown in Table 38. These are based upon a labor requirement estimate of one man-hour per day and a flocculant dosage of 1 ppm of thickener influent.

These estimates are very significant in demonstrating low processing costs for large volume plants. These costs, combined with reduced area requirements, superior water quality control and the positive solids handling which a thickener provides, makes their use in large plants more attractive than a system of large settling lagoons.

Table 36. SUGGESTED OPTIMUM DESIGN PARAMETERS FOR THICKENER OPERATION

1.	Rising Water Velocity - gpm/ft^2	0.2 maximum
2.	Solids Settling Rate - ft/hr	2.0 minimum
3.	Unit area ft^2/T/day	300 minimum (may exceed 10,000)
4.	% Solids in Underflow	5.0 minimum
5.	Sludge Recycle Provision - % Circulating Load	5 to 30
6.	% Solids to Feed Well (Combined raw feed plus recycle sludge)	1 minimum
7.	Flocculant Dosage - ppm influent	1 - 2
8.	Depth of Feed Well Below Sludge Level - ft	2 minimum
9.	Sludge Depth Below Feed Well - ft	6 minimum
10.	Settled Sludge Retention Time - hrs	72 minimum

Table 37. APPROXIMATE THICKENER EQUIPMENT COSTS

CMD plant capacity, gpd	Thickener size diam., ft	Thickener cost, $	Annual depreciation (10 yr basis), $	Daily depreciation $	Capital cost (¢/ 1000 gal)
100,000	25	55,000	5,500	15	15.0
250,000	40	72,000	7,200	20	8.0
500,000	50	80,000	8,000	22	4.4
1,000,000	100	100,000	10,000	27	1.0
5,000,000	150	150,000	15,000	41	0.8
10,000,000	200	250,000	25,000	69	0.7

Figure 26. Sizing and thickener cost estimation data

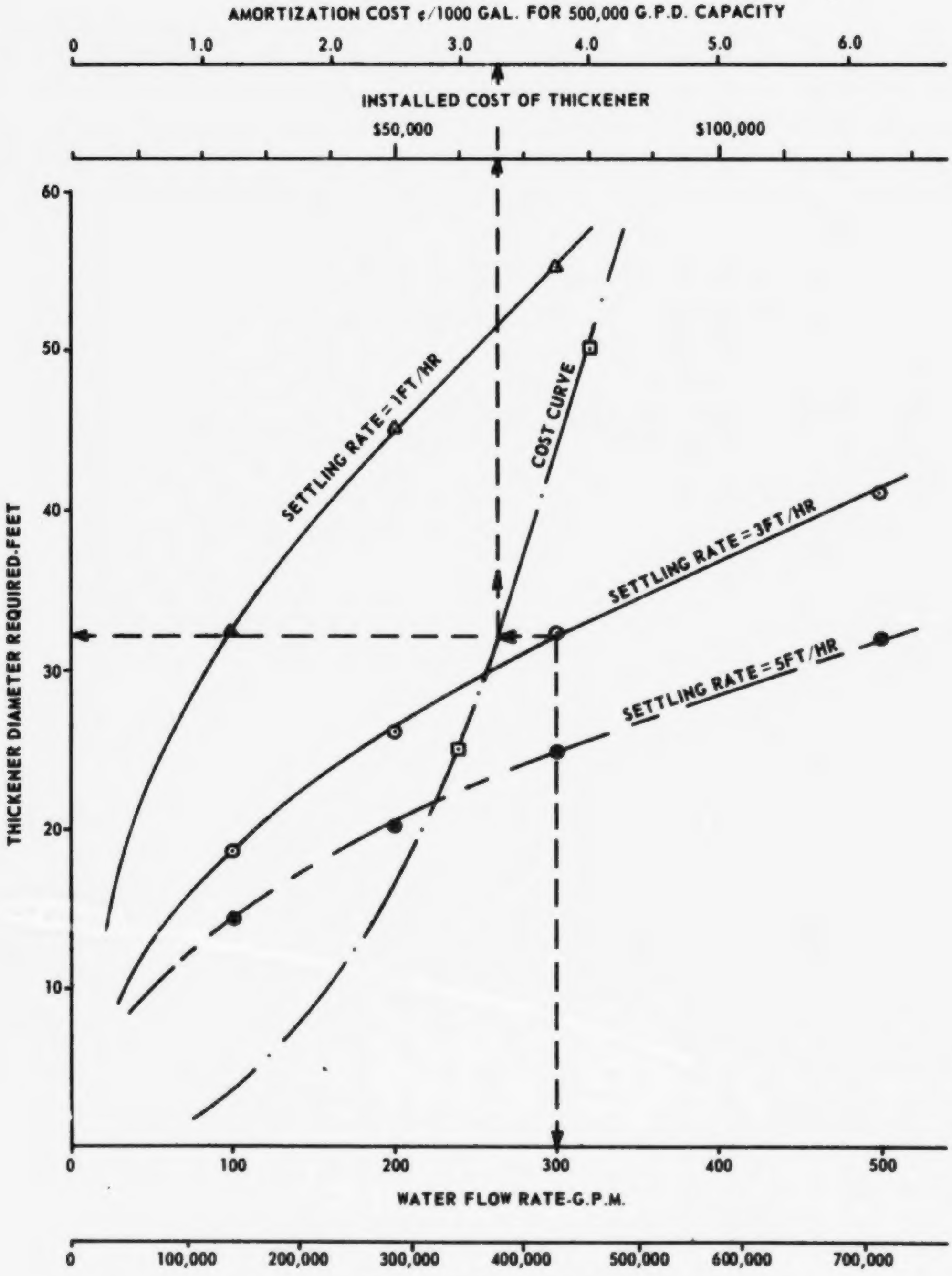


Table 38. DAILY THICKENER OPERATION COSTS

CMD plant capacity - gpd	250,000	1,000,000	5,000,000
Thickener Diameter - ft	25.00	100.00	150.00
Equipment Cost - $	72,000.00	100,000.00	150,000.00
Annual Depreciation (10 year basis) - $	7,200.00	10,000.00	15,000.00
Operating Costs			
Direct Labor - @ $3.50/hr - $	3.50	3.50	7.00
Supervision (25% Direct Labor)	0.88	0.88	1.75
Utilities			
Electrical - @ 1.5¢/kwh - $	1.20	4.80	24.00
Flocculants (1 ppm) - @ $1.25/lb - $	2.63	10.52	52.00
Maintenance (1% equipment cost/yr) - $	1.97	2.74	4.11
Total Operating Cost - $	10.18	22.44	89.46
Amortization - $	19.73	27.40	41.10
Total Unit Operation Cost - $/day	29.91	49.80	130.56
Cost of Thickening - cents/1000 gal CMD	12.01	5.0	2.6

Summary and Conclusions for Thickener Operation--

Pilot plant experience confirmed engineering predictions that a thickener is the key primary dewatering process for any plant involving the treatment of flows greater than 750,000 gpd with a neutralized suspended solids content of 1,000 mg/l or greater. The precipitated solids handling and disposal problems arising from such waters require the superior process control that only a thickener can provide. This is an essential requirement for an effective, low-cost treatment process and is the logical alternative for settling lagoons. The great advantage of a thickener over the settling lagoon is the greater efficiency and the less costly sludge handling by making secondary dewatering feasible.

Settling Lagoons

The objectives of the settling lagoon studies were to observe: lagoon performance during year-round operation as a water clarifying process; flow and turbulence within the lagoon; secondary water quality changes within the lagoon; and the response to atmospheric factors as temperature, precipitation, and wind. Further, information was desired on the nature of sludge accumulation and changes in sludge properties which control its handling by gravity flow and by pumping.

Simplicity, convention and apparent low cost have encouraged the use of this unit operation. However, maintained and reliable performance of lagoons is found to involve many complex functions, all of which are seldom optimized. Among the functions frequently sought, in addition to solid-fluid separation, are: uniform effluent quality based on buffering response of large volumes, providing treated water for other uses, effluent quality monitoring, retention time to complete reactions, and the control to densify, dewater, store, and even dry the sludge. The particular functions required will determine the design. The lagoon system types may be grouped: clarification of low level solids requiring minimum sludge storage (Calhoun[24]), clarifier-thickener providing for temporary sludge storage and intermittant removal of sludge (most common approach), total sludge control without sludge removal (requiring very large multiple lagoons (Mihok[8]), and secondary sludge storage including permanent disposal (Holland[28]). Providing capability for intermittant sludge removal usually limits the lagoon size. The process factors which determine lagoon design are: flow rate, unit sludge volume, slurry settling rate, sludge characteristics, and requirements for secondary dewatering of the sludge. Weather conditions, including heavy rainfall, wind, temperature layering and freezing, can drastically affect operation of a settling lagoon but these effects may be partially counteracted by design.

Lagoon operation based on continual sludge deposition is not only subject to decreasing retention time but currents and particle suspension which lead to particle discharge in the effluent. deVilliers[76] reports on design factors contributing to these difficulties. However, such trends in lagoon design modification increase costs and lead to mechanical sludge removal and peripheral effluent discharge--just a step removed from a thickener.

Basis for the Hollywood Settling Lagoon Design--

Seeking compliance with Sanitary Water Board guidelines[10a] and versatility with minimal construction cost, the unique lagoon design incorporated a bottom, gravity flow sludge removal capability. It is illustrated in Figure 4 and in Appendix B, Figures 4, 5 and 11. Some of its major design features include:

1. Three-compartment, sectionalized construction permitting operation of the influent section as a thickener, with the following sections acting as clarifying and polishing operations.

2. Full lagoon-width influent weir minimizing velocity and providing a uniform feed stream at the surface leading to effective utilization of the entire lagoon. This greatly reduced turbulence and bottom agitation at the feed end. This action and a typical sludge settling pattern is shown in Figure 27.

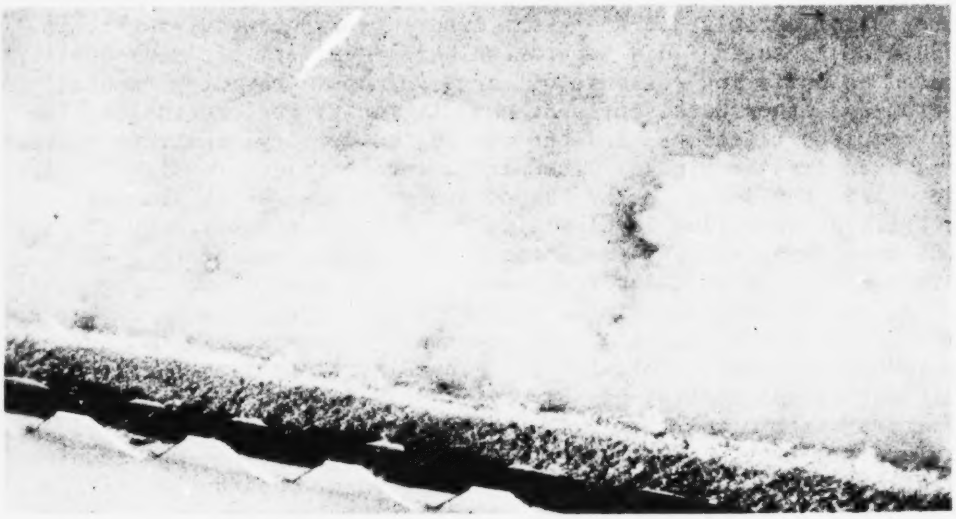


Figure 27. Settling Lagoon Weir Overflow Showing Sludge Settling Pattern.

3. Transverse baffles separating compartments but also minimizing currents, surface windage effects, and short circuiting. The bottom elevation of the baffles defined the design sludge storage volume.

4. Tapered sides and bottom with transverse sludge channels (in first two compartments) allowing bottom sludge removal with minimal disruption of the settling functions or development of turbidity. The slopes assisted the sludge flow to the outlet.

5. Smooth, impervious, bituminous-covered lagoon bottom assisting movement of the settled sludge along slopes.

6. Sludge pump station (Figures 4, 6, 7 and Appendix B, Figures 5 and 10) transferring the settled sludge to secondary dewatering operations.

7. Lagoon effluent structure (Figures 4, 6, 7 and Appendix B, Figures 4, 5, 10 and 11) releasing clear water from surface of lagoon, providing for monitoring flow and quality, permitting transfer of treated water for plant use and allowing variation in water depth.

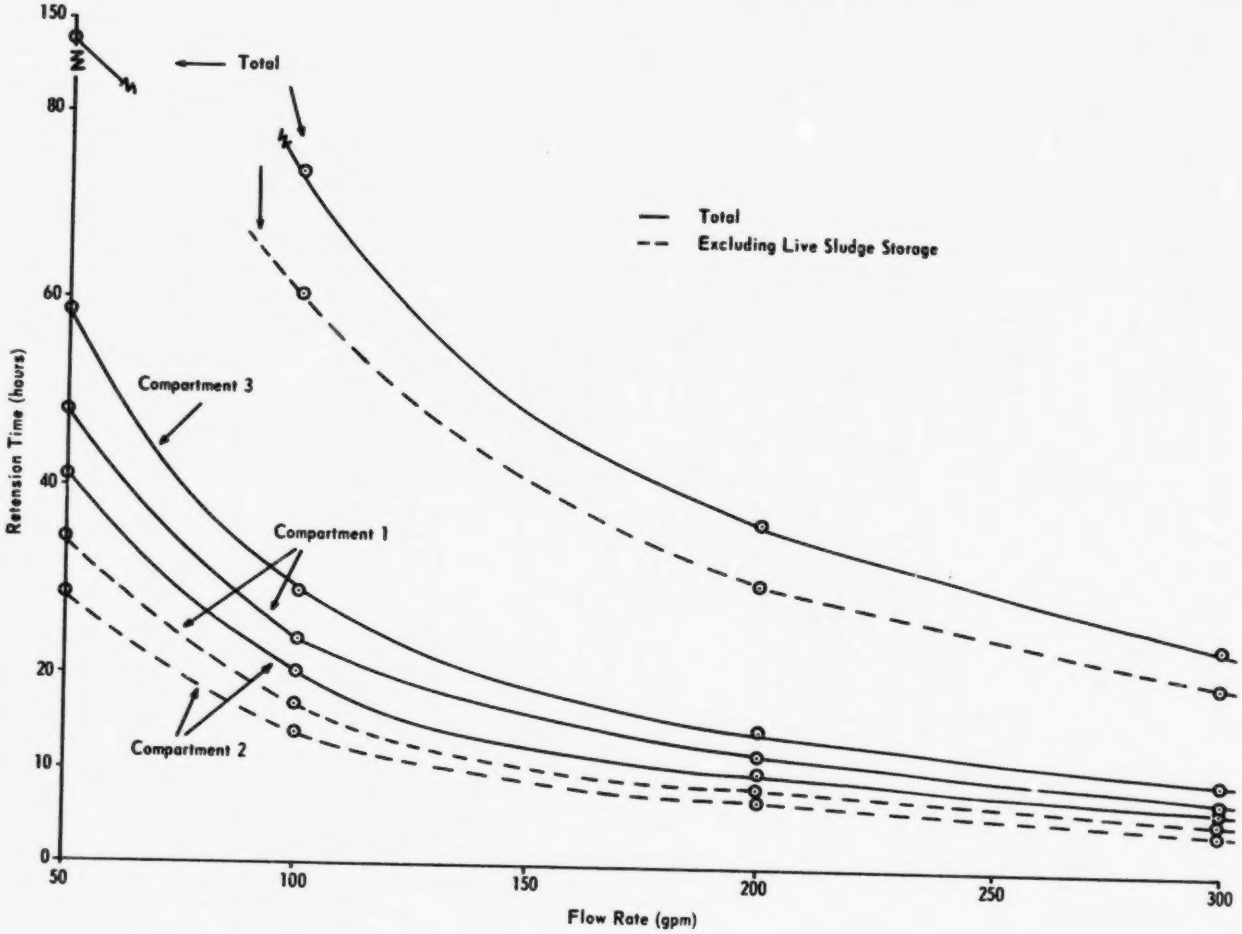
The lagoon retention times at different flow rates are plotted in Figure 28. The changes in retention time with filled sludge storage areas are also shown. Major changes in retention time with flow rate occur at low flows while small retention differences result at high flow rates.

Settling Lagoon Operating Observations--

The settling lagoon was essentially in continuous use during the study as the final settling and water polishing unit operation. This provided observation of long-term operating performance and operating information, especially during cold weather.

Water Handling--Very few water handling problems arose with lagoon except during severe winter weather. Under normal operating conditions, a minimum clear water depth of five feet was maintained in the feed compartment while handling maximum plant flow of 350 gpm containing up to 2,000 ppm of neutralized solids. Retention time was more than adequate even with sludge storage exceeding design capacity. Turbid effluents occurred only when situations unrelated to settling phenomena existed. The baffle and lagoon bottom design controlled short circuiting difficulties. No sludge inversions or floating islands of sludge were experienced. The clarity of lagoon effluents under these conditions was satisfactory for discharge. A major increase in turbidity of the lagoon effluent was associated with

Figure 28. Settling lagoon calculated retention time



heavy rains. Lesser increases in turbidity occurred during rapid sludge removal periods. The heat loss from the quiescent water during severe winter weather permitted ice cake formation on the lagoon surface. Under these conditions, a limited water flow was formed as a narrow channel between the ice, growing down, and the sludge, building up. See illustration in Figure 29. This situation impeded clarification. Under these conditions, the water velocity increased, while the retention time and sludge settling rates decreased. Sludge accumulation was gradually extended into the final compartment and the effluent turbidity increased. Continuous or frequent sludge removal from the bottom minimized this problem, although adequate sludge storage should be provided. If ice prevents surface removal of the sludge, the increased compression, aging, and gelling of the sludge further degrades its flow characteristics.

The adjustable height overflow pipes discharged only the clearest effluent in the lagoon. They were maintained nearly vertical to maximize retention time. There was no apparent advantage from changing their position from center of the effluent structure. The pipes were moved only to lower the water level prior to sludge cleanout operations.

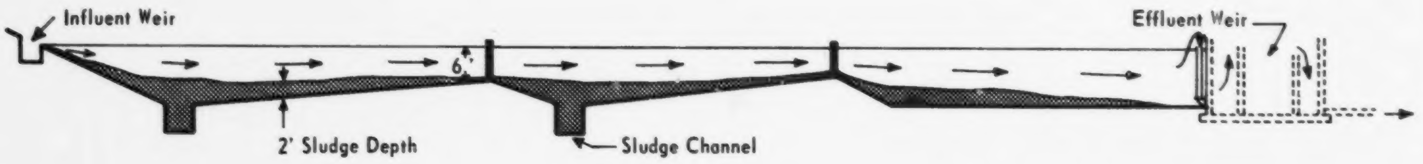
The full-width influent weir compartment was found to be excessively deep which permitted settling of larger or rapid settling particles in the weir. However, the design objective to achieve a low influent velocity and initiate rapid sludge settling functioned well, even during freezing temperatures.

The only effluent clarity difficulties occurred during the use of carbonate reagents or when the preceding neutralization or oxidation reactions were incomplete. The Surface Scatter Turbidimeter evaluating the effluent responded well indicating even minor trends toward an unsatisfactory condition.

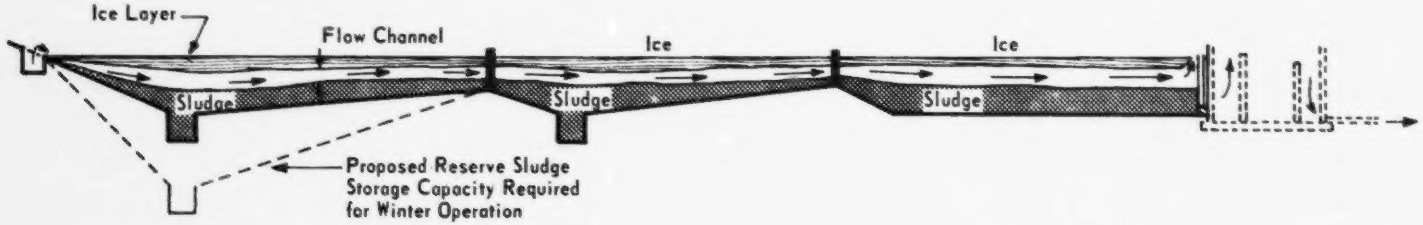
Sludge Handling--Bottom sludge removal from the lagoon was at times troublesome. The observations regarding sludge handling characteristics included:

1. The sludge moved satisfactorily from the bottom channels of the settling lagoon to the sludge drying basin by gravity, prior to its gellation, although it tended to resist flow from the sloped bottoms to the channel. The positive displacement pump in the sludge pumping stations was used only during major lagoon cleaning and in transferring sludge to the Control Building for filtration. The limestone-produced sludge is most difficult to move due to its density. The plant operators were reluctant to transfer sludge from the lagoon to the drying basins at controlled

Figure 29. Sludge deposition in settling lagoon


Influent Weir
6"
2' Sludge Depth
Sludge Channel
Effluent Weir

NORMAL OPERATION


Ice Layer
Flow Channel
Ice
Ice
Sludge
Sludge
Sludge
Proposed Reserve Sludge Storage Capacity Required for Winter Operation

slow rates, seeming to prefer more rapid flows. This psychological reluctance, unexplained, extended beyond the press of other duties. Low velocity flows are preferable since the tendency toward "ratholing" increases with flow rate.

2. The mobility of an "aged", settled sludge, when covered by several feet of water was poor. It would not flow freely from the sloped bottom into the sludge channels and hence through the effluent pipes by gravity flow or from the pump. This indicates that the angle of repose of the submerged sludge is greater than 30°, and slopes approaching 45° are preferable. A slow moving screw with deep blades within the bottom channel would be helpful in directing the sludge toward the exit ports to maintain maximum sludge density.

3. Mobility of aged sludge on the sloped surface of a drained lagoon was satisfactory when movement was initiated by a squeegee or hose water pressure. It tends to move as a large block. On quiescent compression and aging, the sludge forms a gel which has thixotropic tendencies. The gellation behavior develops within 48 hours under sludge and water pressure compression. There was no evidence of this gellation phenomena in the settled sludge in the thickener. Apparently the slow movement of the sludge by the thickener rake prevented this gel formation. A detailed understanding of this aging and gel formation is needed to assist lagoon design and operation. A two- and three-foot thick layer of gelled sludge in one section of the lagoon could be moved into the sludge channel by two men in 4 to 5 hours using squeegees and low pressure hoses. Blocks of sludge were cut with shovels and would slide down the sloping bottom to the sludge channels, provided the bottom was kept thoroughly wetted by a small flow of plant water from a hose. Such settled sludge would normally range between five and 15 percent solids. This procedure was rapid and minimized sludge dilution. There was no evidence that the bituminous bottoms significantly enhanced sludge movement, but they did maintain stable dimensions and resist erosion or mud formation during sludge removal. The side walls of the lagoon were earthen. Although no significant erosion problems developed with the compacted clay lagoon walls, sludge removal from these areas was more difficult. An occasional small stone from the lagoon walls would block the check valve of the sludge pump. Total bituminous side and bottom surfaces are definitely advantageous and recommended. Under no conditions should a stone rip-rap be employed below water levels in a lagoon.

4. Transfer of gelled sludge from the settling lagoon to the drying beds by gravity flow was unsatisfactory. This sludge form tends

to "rat-hole" quickly, allowing excessive amounts of clarified supernatant to move with the sludge and complicate subsequent operations.

5. The sludge had adequate flow characteristics in lines and through pumps when movement began. The sludge transfer lines were flushed with clear water after ceasing sludge transfer to prevent scaling or further gellation. No line blockage was experienced.

6. Sludges produced with hydroxide-type alkalies were the least dense and did not adhere very tightly to the lagoon bottom. In contrast, settled limestone slurries were dense, sticky, and clay-like in character and adhered tenaciously to the bottom surface.

7. The distribution of sludge among the three compartments of the settling lagoon could be related to the sludge settling rates. Sludges settling at rates in excess of 2 ft/hr seldom passed into the second compartment. When the sludge was not removed and the first compartment approached its sludge capacity level (bottom of baffles), the settled sludge was carried into the second compartment. During one three-month period without sludge removal, some sludge collected in the final polishing compartment.

   Slower settling, Type II sludges, tended to drop more uniformly throughout compartments one and two with little deposition in the final compartment. The very rapidly settling sludges, such as those from limestone treatment, accumulated on the sloped bottom of the first compartment, a few feet from the weir. See rapid settling shown in Figure 27. Only as the accumulation in this region increased, did the sludge storage extend beyond the sludge removal channel. The sludge removal practice sought to maintain low sludge storage levels in the first compartment. There were distinct differences in the sludges settled in the three compartments. That settling in the second compartment was less dense and tacky with smaller, lighter floc. These sludges showed lesser tendencies toward gelling, "rat-holing" and were easier to move. They maintained a smaller angle of repose.

Settling Lagoon Performance Data--

The settling lagoon functioned well as a clarifier in dewatering. The turbidity values of the effluent ranged between 1 and 5 jtu with suspended solids between 2 and 13 ppm. The ditch returning the plant effluent to Tyler Run showed no staining after the year's program. Satisfactory water quality was indicated by several generations of animal and plant growth in the third lagoon compartment. These included toads, frogs, salamanders, and insects. Trout and bass transferred to this section of the lagoon found ways of accumulating excessive-levels of sludge in their gills, although they survived well in the treated waters transferred from the lagoon. Deer and bear frequently drank from the lagoon. Operating data for the four water sources and several reagents are given in Tables 39 and 42. The slow settling floc of Type II sludges gave the highest effluent solids level, although not necessarily the highest turbidity. The suspended solids and iron levels of these effluent waters met specifications at the highest flows hydraulically possible through the plant. The lagoon was operated once continuously without sludge removal for a sufficient period of time to consider its "live" sludge storage capacity. The longest service (over a period of three months) was equivalent to 42 days of continuous operating time at an average influent flow rate of 175 gpm (250,000 GPD). Assuming the formation of 12,500 GPD of settled sludge (5 percent of original volume) and a live sludge storage volume of 77,800 gal., the capacity of the lagoon would be only six days of settled sludge production. At the conclusion of long operating period, sludge had accumulated nearly to the influent weir with about three feet of clear water in the first two compartments and about two feet of settled sludge in the third or polishing compartment. The clarity remained satisfactory. This observation suggests that sludge compression had occurred and a live sludge load capacity, based upon 48-hour settling volume, is a conservative but realistic estimate for design purposes. A considerable safety factor beyond the one-month sludge storage recommended by the Pennsylvania guidelines[10a] is achieved. Due to difficulties which can develop during sub-freezing weather, a two-month sludge storage volume seems preferable.

Time limitations of the test program did not permit further consideration of the settling lagoon as a static sludge storage vessel as is commonly being used by many mining companies. It is a practice which should be avoided by more frequent sludge removal.

The continuous or frequent interval removal of sludge from a settling lagoon is definitely advantageous. When sludge was permitted to accumulate in the lagoon and its removal was carried out, many difficulties were experienced. Removal would have been most difficult during freezing winter conditions. Labor requirements to manually assist sludge removal, even with described bottom design, was high. It appears that a settling lagoon is not an optimum unit operation in a mine water treatment flowsheet except for plants with small capacity or when treating waters with low solids loading.

Table 39. TYPICAL SETTLING LAGOON PERFORMANCE - BENNET'S BRANCH WATER

	Pebble Lime	Pebble Lime	Hydrated Lime
Influent			
Flow Rate - gpm	100	200	107
Suspended Solids - ppm	105	95	40
pH	5.4-9.8	5.1-8.3	6.1-6.3
Ferrous Iron - ppm	1-4	1-5	4-6
Alkalinity - ppm	46	14	43
Separated Sludge			
Sludge Volume, %	4.5	4.5	4.5
Settled Sludge Solids, wt. %	0.2	0.2	0.2
Sludge Production, gpd	6,500	13,000	19,800
Dry Solids, lbs/day	108	216	330
Estimated Operational Days[a] without sludge removal			
Compartment No. 1	6	3	2
Compartments No. 1 and 2	12	6	4
Feed Dilution[b]	9,537	10,500	25,100
Final Dilution[b]	500	500	500
Unit Area, ft²/t/day[c]	24,093	26,460	65,582
Effluent			
pH	7.0-8.9	6.6-9.6	6.3
Ferrous Iron - ppm	0-1	1-4	2-4
Suspended Solids - ppm	6	8	7
Temperature - °F	32	32	37
Alkalinity - ppm	16	31	15

[a] Based on lagoon sludge storage capacity and live sludge volume.

[b] Based on liquid to solid weight ratios.

[c] See settling rates in thickener section.

Table 40. TYPICAL SETTLING LAGOON PERFORMANCE - TYLER RUN WATER

	Pebble Lime	Pebble Lime	Hydrated Lime
Influent			
Flow Rate - gpm	100	200	344
Suspended Solids - ppm	70	55	65
pH	8.4-9.6	7.6-12.0	5.6-9.8
Ferrous Iron - ppm	0-1	0-1	-
Alkalinity	120	15	-
Separated Sludge			
Sludge Volume, %	2.0	2.0	2.0
Settled Sludge Solids, wt. %	0.2	0.2	0.2
Sludge Production, gpd	2,880	5,760	9,907
Dry Solids, lbs/day	48	96	165
Estimated Operational Days[a] without sludge removal			
Compartment No. 1	14	7	4
Compartments No. 1 and 2	27	14	8
Feed Dilution[b]	13,330	18,200	15,400
Final Dilution[b]	500	500	500
Unit Area, ft²/t/day[c]	68,408	94,376	79,444
Effluent			
pH	6.5-9.6	8.9-10.6	6.6-7.5
Ferrous Iron - ppm	0-1	0-1	0-1
Suspended Solids - ppm	25	15	18
Temperature - °F	38	42	65
Alkalinity - ppm	10	20	13

[a] Based on lagoon sludge storage capacity and live sludge volume.

[b] Based on liquid to solid weight ratios.

[c] See settling rates in thickener section.

Table 41. TYPICAL SETTLING LAGOON PERFORMANCE - PROCTOR NO. 2 WATER

	Pebble Lime	Pebble Lime	Pulverized Limestone
Influent			
Flow Rate - gpm	100	200	133
Suspended Solids - ppm	1190-5800	2289	1356
pH	7.3-9.0	7.5-10.0	7.5-8.1
Ferrous Iron - ppm	20	0-1	0-1
Alkalinity - ppm	60	50	83
Separated Sludge			
Sludge Volume, %	29.6	29.6	2.0
Settled Sludge Solids, wt. %	0.46	0.46	6.8
Sludge Production, gpd	42,624	85,248	3,830
Dry Solids, lbs/day	1,636	3,273	2,173
Estimated Operational Days[a] without sludge removal			
Compartment No. 1	0.9	0.5	11
Compartments No. 1 and 2	1.8	0.9	20
Feed Dilution[b]	454	456	773
Final Dilution[b]	20	20	15
Unit Area, $ft^2/t/day$[c]	579	581	246
Effluent			
pH	8.0-9.5	9.4	7.2
Ferrous Iron - ppm	0-4	0-1	2
Suspended Solids - ppm	4	3	10
Alkalinity - ppm	20	50	65
Temperature - °F	40	40	45

[a] Based on lagoon sludge storage capacity and live sludge volume.

[b] Based on liquid to solid weight ratios

[c] See settling rates in thickener section.

Table 42. TYPICAL SETTLING LAGOON PERFORMANCE - PROCTOR NO. 1 WATER

	Pebble Lime	Pebble Lime	Hydrated Lime
Influent			
Flow Rate - gpm	100	200	237
Suspended Solids - ppm	-	360	380
pH	5.7-8.4	5.4-9.5	6.3-6.6
Ferrous Iron - ppm	1-11	1-6	1-3
Alkalinity - ppm	99	160	-
Separated Sludge			
Sludge Volume, %	10.0	10.0	8.0
Settled Sludge Solids, wt. %	0.35	0.35	0.10
Sludge Production, gpd	14,400	28,800	27,302
Dry Solids, lbs/day	421	841	228
Estimated Operational Days[a] without sludge removal			
Compartment No. 1	3	1.4	1.5
Compartments No. 1 and 2	5	2.7	2.8
Feed Dilution[b]	2,878	2,878	2,782
Final Dilution[b]	286	286	1,000
Unit Area, $ft^2/t/day$[c]	987	987	505
Effluent			
pH	7.4-9.3	6.8-9.4	6.4-6.5
Ferrous Iron - ppm	13	0-1	2-6
Suspended Solids - ppm	10	5	6
Temperature - °F	35	38	-
Alkalinity - ppm	-	35	45

[a] Based on lagoon sludge storage capacity and live sludge volume.

[b] Based on liquid to solid weight ratios.

[c] See settling rates in thickener section.

Cost of Settling Lagoons--

In addition to sludge settling rates, the sludge storage and/or removal capability must be established for design purposes. Live sludge storage capacity must be adequate for the service time required of the lagoon. This capacity is defined as the 48-hour settled sludge volume. With highly loaded waters, the 48-hour settled sludge volume will fall within the range of 5 to 20 percent of the treated water volume.

Assuming lagoon operation by sludge sedimentation, temporary sludge storage, and interval surface sludge removal, cost estimates for several lagoon sizes are given in Table 44. The plant capacities, sludge development levels, and provision for two-months' sludge storage are developed in Table 43. The costs include the elementary excavation of lagoons 12 feet deep under fair soil conditions. Soil compaction, complex shapes and bituminous-covered surfaces will increase costs. Construction must include compaction and erosion control. The one-month sludge storage capacity recommended[10a] is generally inadequate to insure good operation in subfreezing weather in northern Appalachia, while the five-foot minimal clear water depth may be too conservative. The 48-hour sludge volume basis may be too conservative as Hollywood data and observation by others[28] indicate some sludge volume reduction (up to 50 percent) while under long term compression; however the basis appears to incorporate a highly desirable safety factor.

The simple pond concept used for the estimates represents the inefficient, troublesome units now in widespread use. The lagoon sizes calculated for plants handling greater than one million GPD are larger than those presently in operation in the same flow capacity bracket. Many of these lagoons perform satisfactorily for only a brief period. Difficulties are experienced from charging excessive amounts of suspended solids and in sludge removal. The costs of these lagoons are high. They have minimal versatility and no intrinsic or salvage value. For large lagoons, the land area requirements are significant and costly. Difficulties in sludge removal increase with size. At some locations, existing topographic features, including abandoned surface mines, may reduce these costs.

Comparative sizes and costs for thickeners and simple settling lagoons are shown in Table 45. With 500,000 GPD capacity, the costs are about equal, while for larger systems, equipment costs and space requirements are significantly lower for thickener installations. Since a thickener is the ideal settling lagoon, careful evaluation and comparison between a thickener and settling lagoon should be considered for the primary sludge removal of systems treating large volumes of neutralized coal mine water. Settling lagoons should be

Table 43. SETTLING LAGOON AREA AND VOLUME REQUIREMENTS
Basis: Lagoon providing 10 foot depth of live settled sludge working volume

CMD Plant Capacity gpd	48-hr Settled sludge volume vol. % of raw CMD	Daily volume settled sludge gals.	Area required for 10-foot sludge depth accumulation for			
			One day	60 Days	One year	
			ft^2	ft^2	ft	Acres
100,000	2	2,000	27	1,600	9,600	0.3
	5	5,000	67	4,000	24,000	0.7
	10	10,000	134	8,000	48,000	1.4
250,000	2	5,000	67	4,000	24,000	0.7
	5	12,500	167	10,000	60,000	1.4
	10	25,000	334	20,000	120,000	2.8
500,000	2	10,000	134	8,000	48,000	1.4
	5	25,000	334	20,000	120,000	2.8
	10	50,000	668	40,000	240,000	5.5
1,000,000	2	20,000	267	16,000	96,000	2.8
	5	50,000	668	40,000	240,000	5.5
	10	100,000	1,336	80,000	480,000	11.0
5,000,000	2	100,000	1,336	80,000	480,000	13.8
	5	250,000	3,340	200,000	1,200,000	27.5
	10	500,000	6,680	400,000	2,400,000	55.0
10,000,000	2	200,000	2,672	160,000	960,000	27.5
	5	500,000	6,680	400,000	2,400,000	55.0
	10	1,000,000	13,360	800,000	4,800,000	110.0

Table 44. SETTLING LAGOON EXCAVATION COST ESTIMATES

Settling Lagoon Excavation Cost Estimates

Basis: 10 ft depth of live sludge bed plus 2 ft depth of clear surface water.

Costs: \$1.70/yd^3 (< 100,000 yd^3), \$1.50/yd^3 (100,000 - 1,000,000 yd^3), \$1.00/yd^3 (> 1,000,000 yd^3)

Plant Capacity gpd	Settled Sludge Volume %	Excavation Requirements - for 12' Depth						Excavation Cost \$	
		60-Day Service			1-Year Service				
		Surface Area		Volume	Surface Area		Volume	60-Days Service	1-Year Service
		ft^2	acres	cu. yds	ft^2	acres	cu. yds		
100,000	2	1,600	0.04	710	9,600	0.22	4,250	1,200	7,230
	5	4,000	0.09	1,780	24,000	0.55	10,600	3,020	18,000
	10	8,000	0.18	3,550	48,000	1.10	21,200	6,040	36,000
250,000	2	4,000	0.09	1,780	24,000	0.55	10,600	3,020	18,000
	5	10,000	0.23	4,440	60,000	1.37	26,600	7,550	45,200
	10	20,000	0.46	8,880	120,000	2.75	53,200	15,100	90,400
500,000	2	8,000	0.18	3,550	48,000	1.10	21,200	6,040	36,000
	5	20,000	0.46	8,880	120,000	2.75	53,200	15,100	90,400
	10	40,000	0.92	17,800	240,000	5.50	106,400	30,200	160,000
1,000,000	2	16,000	0.37	7,100	96,000	2.20	42,400	12,000	72,000
	5	40,000	0.92	17,800	240,000	5.50	106,000	30,200	160,000
	10	80,000	1.84	35,600	480,000	11.00	212,000	60,400	320,000
5,000,000	2	80,000	1.84	35,600	480,000	11.00	212,000	60,000	320,000
	5	200,000	4.60	88,800	1,200,000	27.50	532,000	150,000	798,000
	10	400,000	9.20	177,600	2,400,000	55.00	1,064,000	266,000	1,064,000
10,000,000	2	160,000	3.68	71,000	960,000	22.00	424,000	120,000	636,000
	5	400,000	9.20	177,600	2,400,000	55.00	1,064,000	266,000	1,064,000
	10	800,000	18.40	355,200	4,800,000	110.00	2,128,000	532,000	2,128,000

Table 45. COMPARISON OF THICKENER AND SETTLING LAGOON SIZES AND COSTS FOR VARIOUS CMD CAPACITIES

CMD plant capacity mm gpd	Settled volume of sludge %	Thickener required		Settling lagoon for 1 year service (12 ft deep)	
		diam.-ft	cost-$	size acres	cost $
0.1	2	25	55,000	0.22	7,230
	5	--	--	0.55	18,000
0.25	2	40	72,000	0.55	18,000
	5	--	--	1.37	45,200
0.5	2	50	80,000	1.10	36,000
	5	--	--	2.75	90,400
1.0	2	100	100,000	2.20	72,000
	5	--	--	5.50	160,000
5.0	2	150	150,000	11.0	320,000
	5	--	--	27.5	798,000
10.0	2	200	250,00	22.0	636,000
	5	--	--	55.0	1,064,000

considered only for small systems, secondary clarifying units, or temporary emergency application.

Lagoon maintenance should be minimal, involving precautions against undesirable seepage, side-wall erosion, and corrosion of structures from the high-dissolved-solids-content waters.

SECONDARY DEWATERING OF SLUDGES FROM COAL MINE DRAINAGE TREATMENT

The solids content of any feed slurry to a mechanical dewatering operation is important. In general, feasibility increases and costs decrease with increased feed density[74]. The physical and chemical nature of the sludge and its solids content are determined by the reagents and processes which precede the secondary dewatering. Thus any dewatering process evaluation must involve a coordinated process flowsheet.

Continuous vacuum filtration is a principal industrial dewatering unit operation. The filtration tests, all made on thickener underflow products, sought to establish technical and economic feasibilities for settled mine water sludge.

The major objectives were:

1. To develop filtration techniques attaining solids filtration rates greater than 100 lbs./ft^2/24 hours;
2. To establish relative costs for filtration in respect to other dewatering approaches;
3. To show whether filter cake has advantages in the handling and disposal(or recovery) of sludges;
4. To determine variation in the filtration characteristics of settled solids from various mine waters and the alkalies used to neutralize them;
5. To study the effects and costs of chemical additives (primarily flocculants) on filtration rates.

Technical and Economic Feasibility of Filtration

The technical feasibility of filtration with hydrous iron oxide sludges has been demonstrated in the processing of waste pickle liquors in the steel industry[9, 76, 77] and with sewage sludges[73a, 79, 80]. Typical filtration rates for several types of sludges are shown in Table 46.

Important properties of mine water sludges affecting filtration are:

1. The particle size and the zeta potential of the solids.
2. The pH and temperature of the slurries.
3. The floc size, as enhanced by the use of flocculants or degraded by agitation of high-shear pumps.
4. The solids content of the feed slurry.
5. The chemical composition of the solids, including the presence of impurities as excess neutralizing agents.

Filtration operational techniques and economics must be related to subsequent handling and disposal of the sludge filter cake and the filtrate. Among these factors are:

Table 46. FILTRATION RATES FOR SEVERAL VARIETIES OF SLUDGES

Sludge	Reagent used	% Solids in feed	Filtration Rates		Type of Filter
			Liquid gpm/ft^2	Solid lb/ft^2/24 hrs	
Mine Water Sludge[41]	Limestone	1.2	0.12	120	Rotary Drum String Discharge
Pickle Liquor[9]	Hydrated Lime (carbide sludge)	≈50	-	>500	Disk Filter
Pickle Liquor[78]	High Calcium Lime	40	0.75-1.13	2600-3800	Pre-coat Test Leaf
Pickle Liquor[77]	Limestone	5	-	110	Vacuum Drum
Raw Sewage Sludge[79]	-	10	-	240	Vacuum Drum
Raw Sewage Sludge[80]	$FeCl_3$-CaO	-	-	89-240	Vacuum Drum
Sewage Sludge[73a]	-	-	-	25-250	Vacuum Drum

1. Clarity of the filtrate.

2. Moisture content of the filter cake, which will determine its bulk handling characteristics, the equipment required for transferring, costs of transportation and ultimate disposal.

Plant Operation - Equipment Used--

The filtration tests used a precoat rotary drum filter with diatomaceous earth (Celite 545) as a filter aid. The EIMCO filter was 6 feet in diameter by 6 feet in length, having 113 ft^2 of filter area (see Appendix B, Figure 9, and its operation in Figures 30 and 31). The filter is of acid-proof construction and has a variable speed drum drive, a variable speed precoat knife-cutting mechanism, a variable rate slurry feed pump, and a slurry flocculant conditioning system.

Bench tests utilized a 0.1 ft^2 test leaf. Some leaf tests were made with strongly flocculated slurries but no precoat with quite promising results. These tests could not be confirmed with the continuous filter because internal filter modifications would have been necessary for its conversion to a non-precoat configuration permitting a blow-back cake discharge.

Filtration - Operating Parameter Limits--

The parameter operational ranges of the filtration unit are listed in Table 47. The values shown under "Maximum Operation Range Experienced" indicate the parameter spread observed while the "Typical Range" values could be maintained under steady-state condition without mechanical operating difficulties.

Filtration - Bench Scale Filtration Testing--

Sufficiently large volumes of thickener sludge were not available to make continuous filtration tests on all types of neutralized sludges produced. In providing the desired additional, supplementary data, a bench scale vacuum filtration test station was operated.

The conventional equipment used in the bench testing is shown in Figure 32. The 0.1 ft^2 EIMCO test leaf was covered with the same filter cloth as the large filter and was provided with a one inch deep brass guard ring which permitted the deposition of about 1/2-inch of Celite 545 on the filter cloth. Excess Celite was removed to provide a precoat surface level with the guard ring thus permitting the filter cake to be removed.

The filtration cycle was varied. The filter cake was carefully lifted from the filter medium with minimum medium contamination.

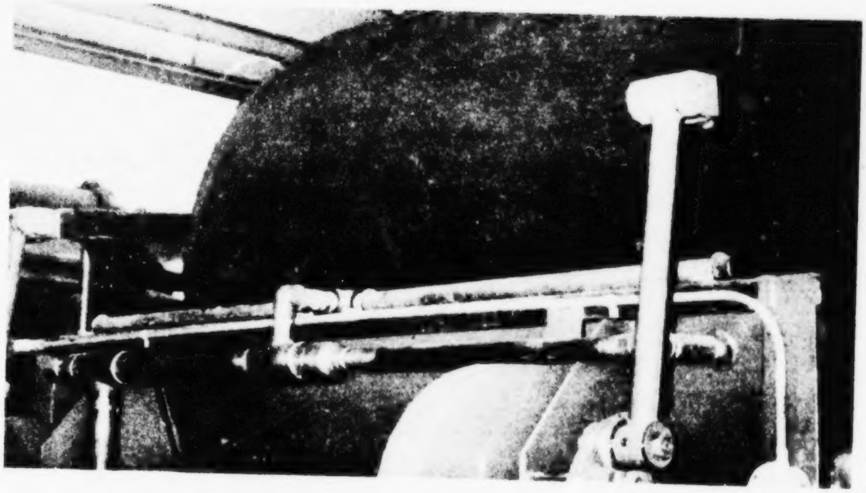


Figure 30. Mine Water Sludge Coating on Precoat Drum Filter

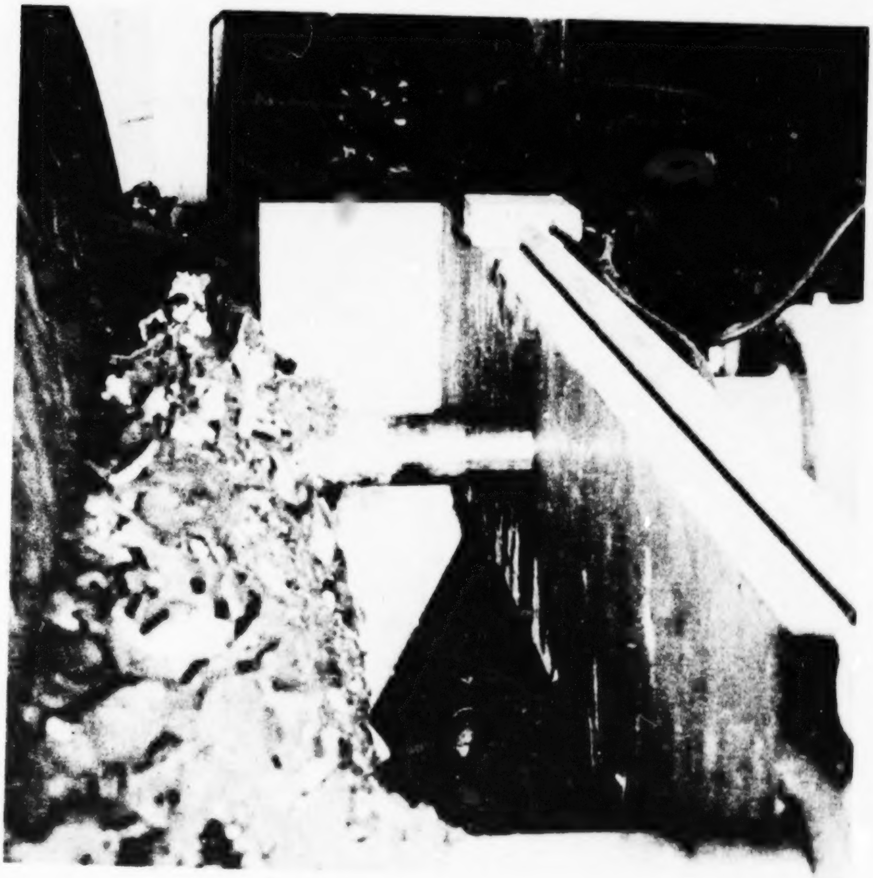


Figure 31. Cutting Sludge Cake from Vacuum Filter

Table 47. OPERATIONAL PARAMETERS FOR FILTRATION UNIT OPERATION HOLLYWOOD PILOT PLANT

Parameter	Maximum operating range experienced	Typical range
1. Filter Slurry Feed Rate - gpm	5-40	5-20
2. Percent Solids in Feed	0.8-15	1-3
3. Dry Solids Feed Rate - lbs/min	0.35-50	1.26-4.7
4. Filter Drum Speed - rpm	0.22-0.85	0.29-0.73
5. % Submergence	18-23	20-23
6. Vacuum - in. Hg.	5-25	18-22
7. Dry Solids Filtration Rate[a] (lbs/ft^2 filter area/24 hours)	10-300	10-25
8. Liquid Filtration Rate - gpm/ft^2	0.1-0.5	0.1-0.35
9. Temperature of Feed - °F	40-75	60-70
10. Percent Solids in Filter Cake[b]	15-27	18-24
11. Filter Cake Thickness - inches	1/32-1	1/16-1/4
12. Flocculant Dosage[c] - lbs/ton solids (Atlasep 1A1)	0-9.9	0
13. Filter Aid Used - lbs/hr (Celite 545)	---	50-65
14. Knife Advance Rate - mils/rev.	0.6-7.2	4-5
15. Filter Cloth Used - Eimco No. PO-801 HF - polyethylene monofilament 1 x 1 plain weave - 112 x 48 thread count per sq. in. - 11 oz. per sq. yd. - air flow = 30 cfm/ft^2.		

NOTES:

[a] These rates are for mine drainage sludge only, and do not include weight of the Celite filter aid. Weight percent of Celite in the filter cake ranges from 10 to 40%.

[b] Filter cake is a combination of mine drainage sludge and Celite filter aid. Solids in Celite 545 filter medium wet cake run about 43%.

[c] A small amount of flocculant was present in most filter feeds, having been added to the thickener at dosages of 1 to 3 ppm of neutralized water.

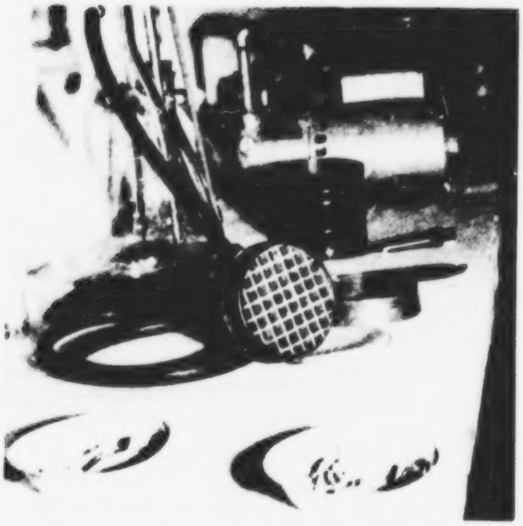
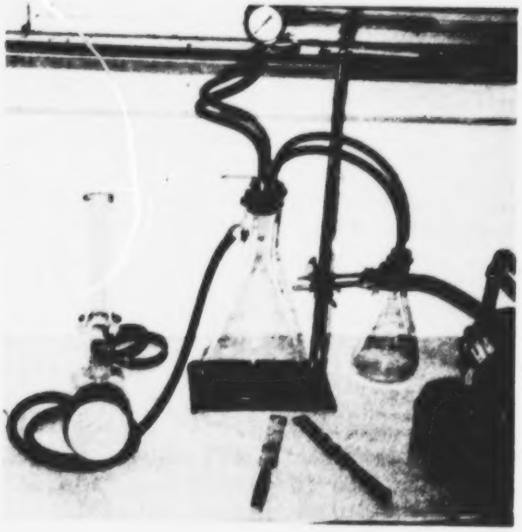


Figure 33. Test leaf filter for top loading process

Figure 32. Test leaf filtration equipment



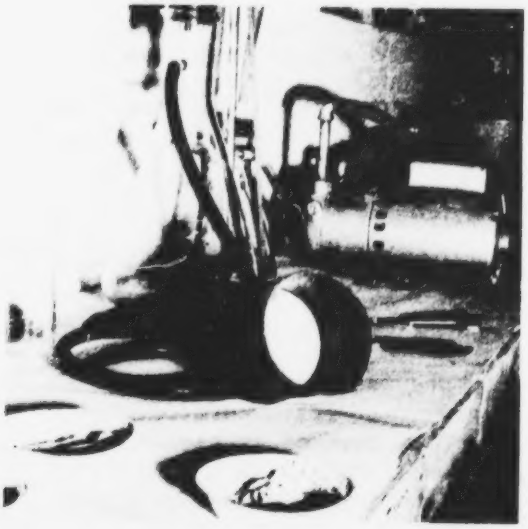


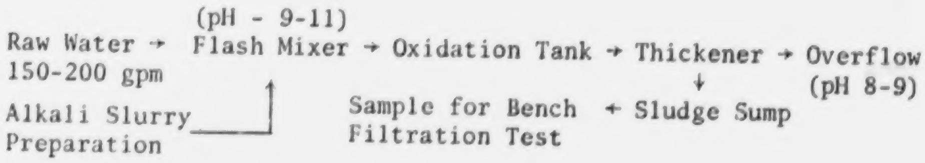
Figure 34. Sleeve for top loading filtration

The wet cake was weighed immediately, dried in an oven at 105°C overnight, and then weighed to evaluate its moisture content. The filter medium remaining on the dried sludge cake was brushed off. This separation can be made easily, permitting the weighing of the sludge directly and computing of the actual filtration rate. A standard filtration test data sheet showing the method of handling the data is given in Table 48.

Bench Scale Filtration Test Data--

Cycle Time Selection--The filtration cycle sequences were based upon a range of values related to the plant filter at various operating speeds. These cycle sequences are shown in Table 49. Tests made to determine the effect of filter drum speed on solids filtration rate utilized hydroxide neutralized sludge from two waters at three levels of solids content. The results are summarized graphically in Figure 35. Theoretical curves were constructed to illustrate the relationship of liquid filtration flow rate to solids content of the filter feed at two levels of constant solids filtration rates (25.2 and 120 lbs/ft^2/24 hours). These solids rate values illustrate a spread of filter capacities from the economically unattractive to feasible. These theoretical curves are shown in Figures 36 and 37. The liquid filtration rate (i.e., rate of water permeability through the sludge filter cake) will be the chief factor in determining filter capacity. The range of liquid filtration rates which could be attained under reasonable operating conditions, and the corresponding rate of solids build-up were determined experimentally.

Bench Scale Sludge Filtration Tests--A series of sludges were prepared in the plant utilizing different waters and neutralizing reagents. The settled sludges were then subjected to filter leaf and drainage rate tests. They were produced in the plant using the following flow sheet:



Before each run the system was drained and washed thoroughly to insure a representative sludge. The treatment tests were continued under uniform conditions to develop several feet of settled sludge in the thickener. These runs did not permit attainment of an equilibrium condition in the thickener with continuous blow down. However, sludge densities attained were typical. Sludge sample densities for testing were controlled when necessary by additional settling.

Table 48. FILTRATION TEST DATA

Sludge Source: Proctor No. 2 from Densator Alkali Used:				Vacuum: 23 Hg % Solids in Feed: 7.19 Filter Aid: Celite 545 (1/2")						Test Leaf Area: 0.1 ft^2 Temperature: 70°F Flocculant Used: Atlas 1A1 (1.5 ppm in Densator)			

Test no.	Drum speed setting^a	Filtrate volume ml	Filtered Solids Weights - Grams							Solids in wet cake-%	Filtration rates		
			Pan + wet cake	Pan + dry cake	Pan	Wet cake	Dry cake	Filter aid	Dry sludge		solids		Liquid
											lb/ft^2/hr	lb/ft^2/24hrs	GPM/ft
		F	G	H	I	J G-I	K H-I	L	M K-L	N 100K/J	O ExM/45.4	P 24x0	Q 0.16F/A
1	1	100	34.4	11.77	4.57	29.83	7.20	2.40	4.81	24.1	1.26	30.2	0.053
2	1	105	32.0	11.20	4.58	27.42	6.62	1.77	4.87	24.1	1.27	30.5	0.055
3	4	75	22.7	8.96	4.55	18.15	4.41	1.25	3.13	24.3	2.34	56.1	0.13
4	4	75	23.2	8.98	4.55	18.65	4.33	1.18	3.00	23.2	2.24	53.7	0.13
5	4.3	65	20.15	8.32	4.60	15.55	3.72	1.15	2.57	24.0	2.57	61.6	0.134
6	4.3	56	22.04	8.66	4.73	17.31	3.93	1.27	2.63	22.7	2.63	63.0	0.115

[a] See Table 49 for Filtration Time Cycle.

Date: May 25, 1971

Table 49. HOLLYWOOD PLANT VACUUM FILTER CYCLE CALIBRATIONS

Drum Speed Setting	Submergence Time		Drying Time		Discharge Time		Total Cycle Time-sec.	Cycles per Hour	Drum Speed	
	sec.	%	sec.	%	sec.	%			RPM	RPH
	A=ca		B=ab		C=bc		D	E 3600/D		
1	60	20	192	63	52	17	304	11.85	0.20	12
4	23	25	56	61	13	14	92	33.91	0.65	34
4.3	16	21	50	64	12	15	78	46.10	0.77	46

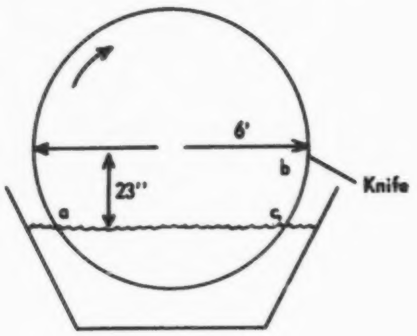


Figure 35. Solids filtration rate as a function of drum speed

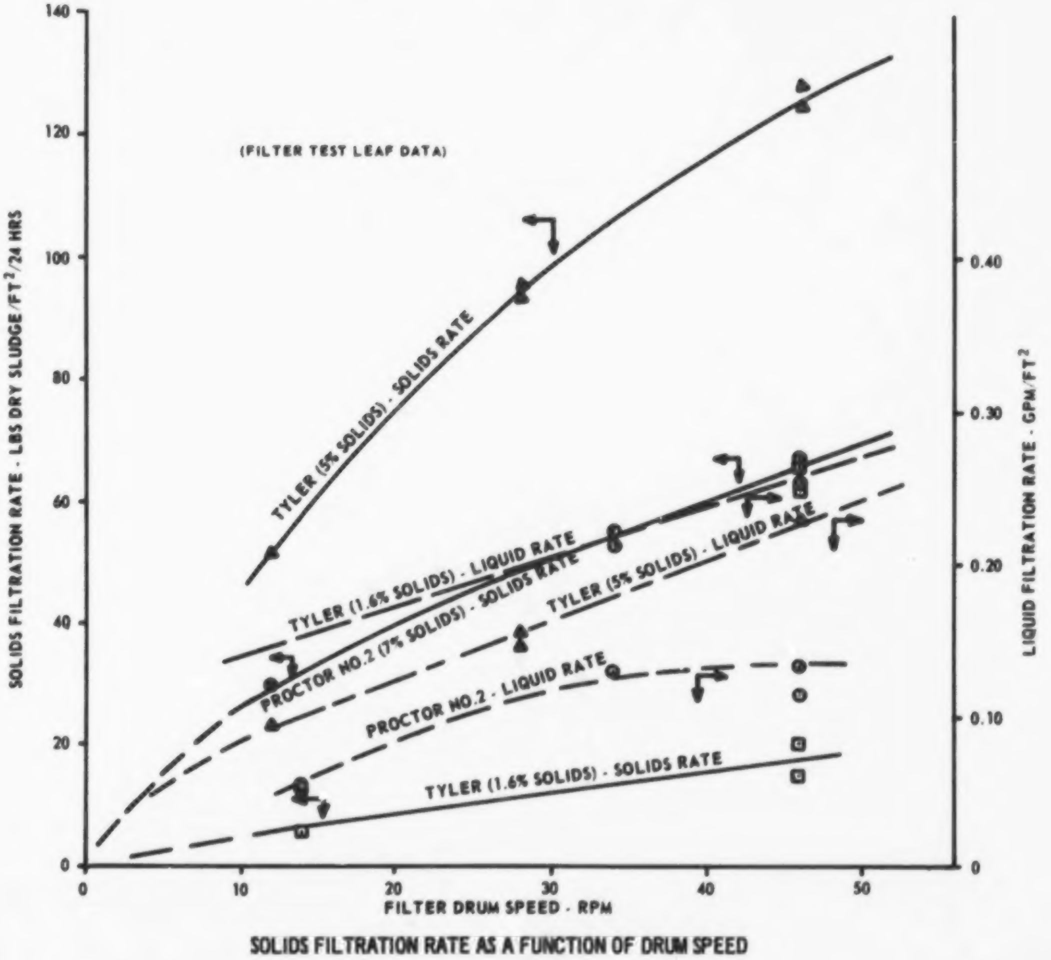


Figure 36. Variation of filtrate flow rate vs. % solids in filter feed at constant solids filtration rate

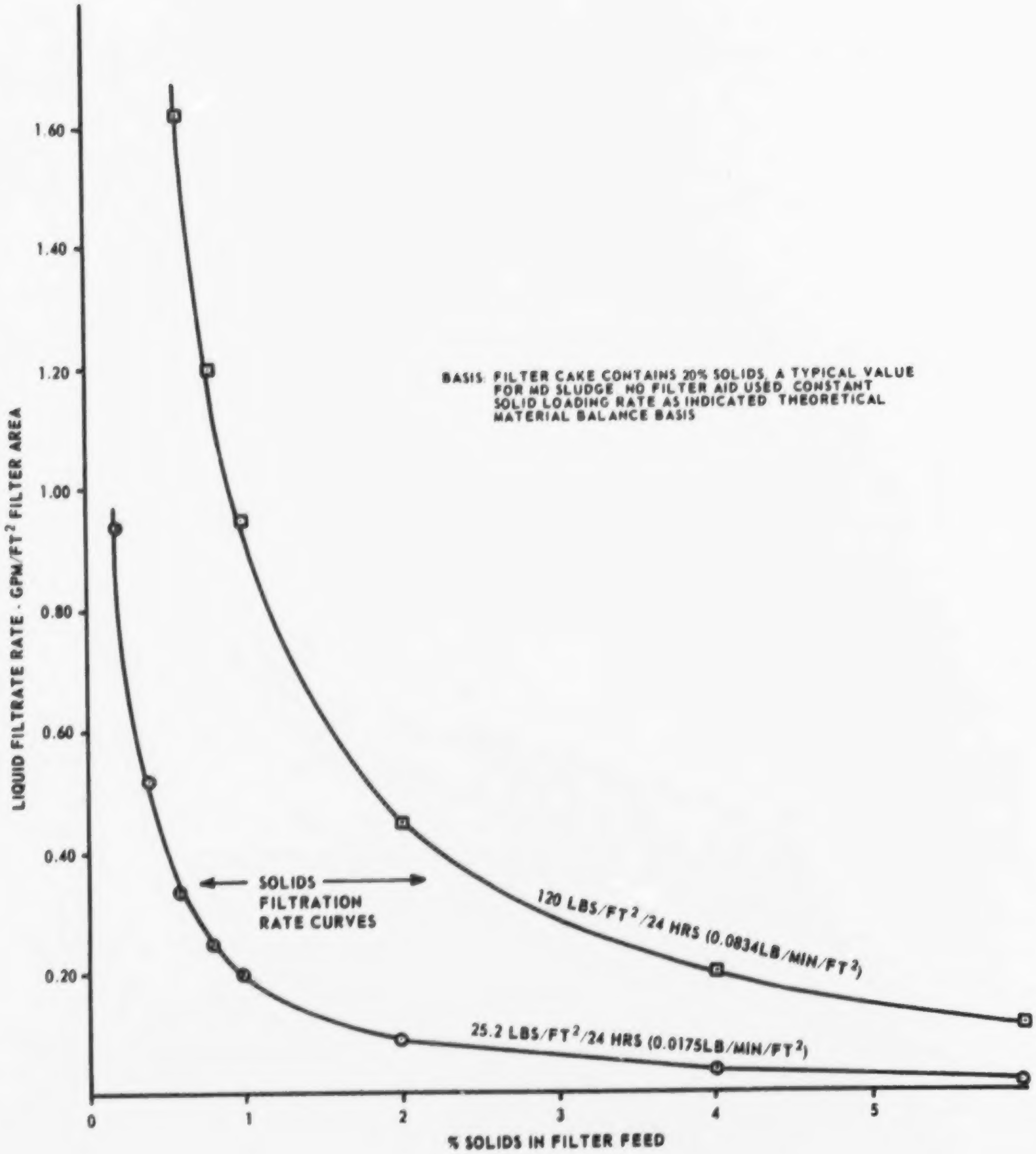
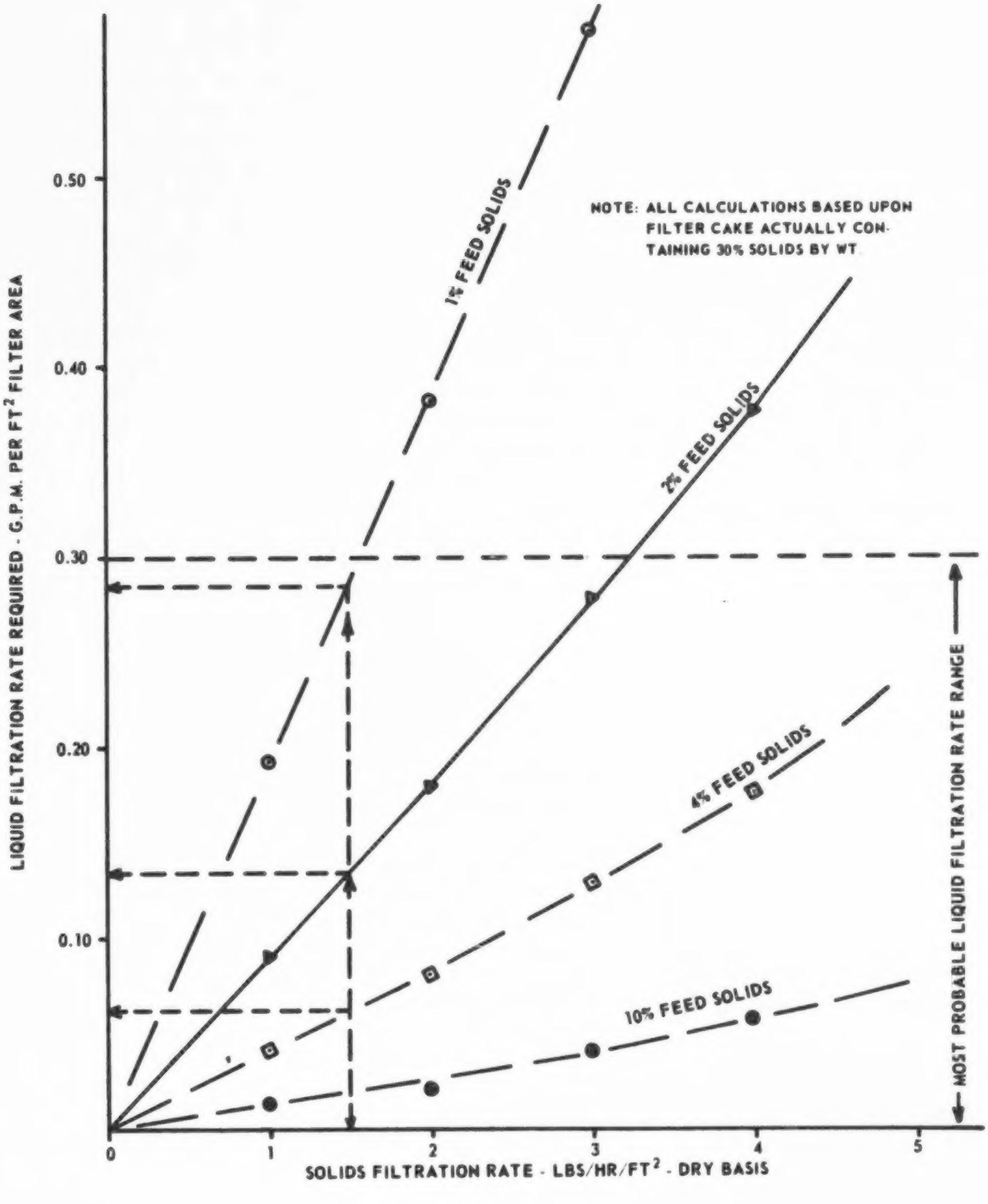


Figure 37. Relationship of corresponding solids and liquid filtration rates at various levels of solids content in the filter feed slurry



The chemical analyses of the sludges are shown in Table 50. In calculating the compound composition, the loss in weight at 180°C was assumed to include waters of hydration of gypsum and basic magnesium carbonate but not from decomposition of metal hydroxides. The sodium content was less than 0.5 percent and not included. The probable compound analyses were based on solubilities of the most likely-occurring compound form in sludges produced at pH 7.0-8.5 and exposed wet to the atmosphere for several days which would convert excess hydroxides to carbonates. All of the sulfate was assumed to be gypsum ($CaSO_4 \cdot 2H_2O$). The calcium in excess of that related stoichiometrically to the sulfate was expressed as $CaCO_3$, since any $Ca(OH)_2$ would have been converted by carbon dioxide absorption. The iron and aluminum were considered to be hydroxides which do not decompose under 300°C. Further, based on solubility and formation conditions, basic magnesium carbonate was assumed to be present. Its reported structure ($5MgO \cdot 4CO_2 \cdot 6H_2O$) is that of the natural mineral, hydromagnesite ($3MgCO_3 \cdot Mg(OH)_2 \cdot 3H_2O$). The solubility of this compound at 25°C in pure water is 0.25 g/l. This compound loses one mole of H_2O at 100°C. The exact form is uncertain but this basic carbonate formula appears to give the most reasonable total balance. Other sludge compositions have been indicated[64].

Summary of Filtration Test Data

The thickener-settled sludges were subject to the filter test leaf, utilizing the previously described procedure and the following filter time cycle.

Submergence	16 seconds	21%
Drying Time	50 seconds	64%
Discharge Time	12 seconds	15%
Total Cycle Time	78 seconds	100%

Cycles/hr - 46.1
Corresponding Filter Drum Speed - 0.77 rpm.

The resulting filter test leaf data are presented in Tables 51 and 52. Only limestone and the aged lime containing $CaCO_3$ showed a major difference in the solids filtration rates or chemical composition. This confirms evidence from many sources (both coal mine water and pickling liquor reports) that calcium carbonate produces sludges with superior settling and filtration characteristics.

Filter feed slurries containing less than two percent solids consistently produce the lowest solids filtration rates (in the range of 15 to 30 $lbs/ft^2/24hrs$). However, at five percent solids or higher, filtration rates of 125 to 331 $lbs/ft^2/24$ hrs were obtained, which makes vacuum filtration a reasonable cost unit operation.

Table 50. CHEMICAL ANALYSES OF SLUDGES USED IN FILTRATION AND DRAINOMETER STUDIES

Weight, % (dry basis 105°-24 hrs)

Alkali Used	Hydrated Lime[a] - Air Oxidation			Hydrated Lime Bio-Oxidation	Hydrated Dolomite	Calcined Dolomite		
Mine Water	Bennett's Branch	Proctor 1	Proctor 2	Proctor 2	Proctor 2	Tyler	Proctor 1	Proctor 2
Component								
Al	3.8	4.7	3.1	8.0	2.8	5.5	4.5	4.8
Fe	19.5	17.7	23.1	24.3	13.0	7.4	13.5	23.2
Ca	6.9	5.8	5.2	4.8	17.2	10.7	6.7	5.2
Mg	6.6	4.3	5.1	1.3	3.8	11.8	9.8	5.8
SO_4	5.7	6.8	5.8	11.5	4.4	1.6	2.3	5.5
H_2O at 180°C	12.5	15.8	14.8	--	10.2	8.7	11.7	14.7
				Compound Composition				
$Al(OH)_3$	11.1	13.7	8.9	23.1	8.2	16.0	13.2	13.9
$Fe(OH)_3$	37.6	34.0	44.3	46.6	24.9	14.2	25.9	44.7
$CaCO_3$	11.4	7.5	7.0	0.0	38.3	25.0	14.4	7.2
$MgCO_3$	--	--	--	--	--	21.0	12.0	6.1
$3MgCO_3 \cdot Mg(OH)_2 \cdot 3H_2O$	25.2	16.4	19.4	5.1	14.4	22.4	24.6	15.6
$CaSO_4 \cdot 2H_2O$	10.7	12.7	10.9	20.7	8.4	3.1	4.4	10.4
Total	96.0	84.3	90.5	95.5	94.2	101.7	94.5	97.9

[a] Warner Company - fresh batch

Table 51. MINE WATER NEUTRALIZED SLUDGE SOLIDS FILTRATION RATES

Mine water source	% Solids in filter feed slurry	Fresh hydrated lime $Ca(OH)_2$	Old hydrated lime[a] $Ca(OH)_2$	Pebble dolomitic lime $CaO \cdot MgO$	Dolomitic hydrated lime $Ca(OH)_2 \cdot MgO$
		Sludge filtration rates - lbs dry solids/ft^2/24 hrs			
Proctor No. 1	1.62	26.4[b] 27.8			
Proctor No. 1	2.53			54.0	
Proctor No. 2	2.70	62.7 72.0			53.2 53.8
Proctor No. 2	7.19		131.0 137.0	63.0 64.0	
Proctor No. 2 (bio)	1.48		39.6 50.3		
Proctor No. 2 (bio)	3.26		82.5 81.4		
Proctor No. 2 (bio)	7.18		331.0		
Bennett's Branch	1.96	33.1 38.6[c]			
Bennett's Branch	2.72	40.1 42.9			
Tyler Run	1.58			15.6 21.2	
Tyler Run	5.03			128.5 125.3	

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[a] In this case, hydrated lime stored in the plant lime storage silo for a period of 3 months, during which time it converted to about 50% $CaCO_3$.

[b] The double values reported for most tests represent duplicate determinations made on each sludge slurry.

[c] No precoat material employed.

Table 52. VACUUM FILTER LEAF TESTS OF SLUDGE FROM LIMESTONE NEUTRALIZED COAL MINE DRAINAGE

Temperature = 70°F

Filter Aid: Celite 545

Sludge Source: Settling Lagoon

Flocculant Used: ATLASEP 1A1 2-3ppm in original neutralized slurry

Water Source: Proctor No. 2 Biochemically Oxidized

Test No.	Vacuum inches Hg	Cycles per hr.	Filtration Rates Solids lbs/ft^2/ 24 hrs	Filtration Rates Liquid GPM/ ft^2	% Moisture in Filter Cake	Filter Aid Condition	Filter Cake Thickness inches
Waters neutralized with Centre County, Pa. limestone (Neidigh Brothers Quarry). Free Solids: 13% by weight							
1	24	46	381	0.16	66.0	Normal Precoat	3/16
2	24	46	451	0.18	65.6	Normal Precoat	3/16
3	16	46	352	0.12	67.0	Normal Precoat	3/16
4	16	46	331	0.12	68.8	Normal Precoat	3/16
5	24	46	394	0.15	65.3	Normal Plus 0.1% Body Feed	3/16
6	24	46	372	0.14	65.6	Normal	3/16
7	24	46	340	0.14	71.0	None	3/16
8	24	23	297	0.21	71.0	None-Topload	1/4
9	24	46	796	0.42	81.0	None-Topload	3/4
10	24	46	995	0.34	78.6	None-Topload	3/4
Waters neutralized with York County, Pa. limestone (Thomasville Limestone Company). Feed Solids: 27.2% by weight.							
11	24	46	990	0.16	55.5	Normal Precoat	3/16
12	24	46	1,200	0.16	51.5	Normal Precoat	3/16

Most of the sludge from limestone neutralization resulted from a flowsheet involving biochemical oxidation and the rotary limestone reactor (see Figure 16, Appendix B), but some had been reacted with pulverized stone in the Densator. Such sludges have rapid settling rates, high solids content in thickener underflows, and excellent filtration characteristics. The filter leaf data given in Table 52 clearly show the excellent filtration rates attainable with these slurries.

The replicate tests using normal precoat vacuum filtration at vacuum levels between 16 and 24 inches Hg yielded greater than 350 lbs/ft^2/24 hrs. Rates in excess of 1000 lbs/ft^2/24 hrs were measured with the higher solids slurry from Thomasville stone. The moisture content of 66 percent is a great improvement over the more typical 80 percent observed with hydroxide slurries. Test seven suggests precoat media may not be necessary. With slurries of this nature, filtration becomes a realistic low-cost method for handling mine water solids.

Filter leaf tests eight, nine, and 10 were made utilizing top loading with a more open cloth and no filter aid was used. Even higher filtration rates were observed by this technique as shown in Table 52.

In this procedure, the vacuum primarily dries the cake and is not needed to load the media. The gravity force will aid rather than hinder the operation. In making these tests a sleeve was added to the 0.1 ft^2 test leaf to permit a deep cake to accumulate as shown in Figures 33 and 34. A three- to four-fold increase in solids filtration rate over submerged-type loading resulted. These tests suggest a top-loading type filter would be most effective.

With flocculated slurries, some individual flocs were so large and cohesive they appeared filterable directly on a filter cloth without the use of precoat. Several tests (Table 57) were made and confirmed this possibility. The results indicate that strongly flocculated pulps do not require filter aid and produce filtrates with good clarity. The elimination of filter aid would remove the highest operating filtration cost factor.

The response on filtration of dilute slurries to a body feed of filter media (one percent of Celite 545) is reported in Tests 20 and 21 of Table 56. This technique greatly improved the filtration of this 1.96 percent solids slurry.

Filtration of High Density Recycle Sludge

In the start up operations of the plant filter, the heavy (> 15 percent solids) recycle sludge was selected anticipating good filtration properties. The response was excellent, in fact the best observed, attaining filter cakes with thickness in excess of 3/4-inches. The cake was dry and readily handleable. The filtration rates were so high that the total Densator storage supply of recycle sludge was consumed in the qualitative start-up procedures. It could not be replaced during the available test period.

Limited data were collected from filter leaf tests with a less adequate, specially prepared sludge. The "limited recycle" sludge was prepared, batch-wise, with gentle agitation in a 230 gallon reactor. Beginning with 50 gallon Proctor No. 2 water, four pounds of aged, hydrate lime was used to raise the pH to at least 12. Fresh drainage was added to lower the pH to 9 with 90 minutes reaction time. After settling and decantation, the procedure was repeated through six cycles.

The final settled solids were filtered and the data summarized in Table 53 was developed. A solids filtration rate in excess of 130 lbs/ft^2/24 hrs offers confirmation of the qualitative tests.

Table 53. FILTRATION RATE TEST DATA FOR RECYCLE SLUDGE

Solids (%) in Settled Sludge	5.0
Solids Filtration Rate	131.0
(triplicate determinations)	138.0
lbs/ft^2/24 hrs	134.0
Liquid Filtration Rate	0.17
(triplicate) gpm/ft^2	0.19
	0.18
Analysis of Raw Water Used	
Total Acidity, ppm	1722
Ferrous Iron, ppm	378
Total Iron, ppm	391
Temperature of Sludge	70°F

The filter leaf test data of settled sludges permit the following conclusions:

1. A realistic operational liquid filtration rate range is 0.1 to 0.3 GPM/ft^2.

2. The highest liquid rates are obtained at the fastest drum speeds (i.e., shortest submergence time).

3. The highest solid filtration rates are obtained at the fastest drum speeds.

4. The highest solids rates are obtained from feed slurries having the higher solids contents and may exceed 1000 $lbs/ft^2/24$ hrs.

5. There is an excellent correlation of the 1.6 percent and 5 percent solids slurries with the theoretical curves in Figure 37, but the 7 percent slurry curve in Figure 35 does not conform.

6. Solids filtration rates of 120 $lbs/ft^2/24$ hrs can be attained without the use of flocculants when using filter feeds containing 5 percent solids. Assuming a limiting liquid rate of 0.2 GPM/ft^2, the theoretical curve shows that a solid rate of 120 $lbs/ft^2/24$ hrs is possible using a 4 percent solids feed.

7. Top loading filter techniques appear to be advantageous.

8. Media body feed is helpful for dilute slurries.

9. Under some conditions, filter media may not be necessary for this type of sludge dewatering.

Pilot Plant Filtration Testing

Filtration testing with the Eimco drum filter was carried out with slurries pumped from the settling lagoon or from the thickener underflow. In many cases, insufficient sludge at constant solids content was available to permit smooth, steady-state operation. The plant data are presented in Table 54. The range of liquid and solids filtration rates are similar to those established in the test leaf filter study. Under normal, full-scale operation, filtration rates greater than 100 $lbs/ft^2/24$ hrs should be possible with a properly designed flowsheet. The chemical composition of these filter cakes is presented in Table 55.

Man-Power Requirements in Filtration

The frequent attention required by the cake cutting knife mechanism is a major labor consideration. Frequent adjustment of the knife was necessary to maintain a clean surface (thus maximum solids capacity) due to changes in compaction of the 3- to 4-inch starting precoat layer. Although continuous operation with a uniform feed minimized this variable, it appears cost estimates should include a significant allowance for operating labor. Taking all phases of this operation into consideration (e.g. flocculant preparation, precoat slurry preparation and application, knife-cutting adjustment, and filter cake disposal supervision), a manpower allowance of about 1/3 man per shift appears realistic.

Table 54. SUMMARY OF TESTS ON PILOT PLANT ROTARY VACUUM FILTER

Vacuum: 22 inches Hg   Temperature: 65° F

Filter Aid: 3-4 inch layer Celite 545   Drum Speed: 46 RPH

Test No.	Sludge Source[b]	% Solids in Filter Feed	% Moisture in Filter Cake	Filtration Rates Solids lbs/ft²/24 hrs	Filtration Rates Liquid GPM/ft²
1	Lagoon	4-6[c]	83.7	6.2[a]	0.07
2	Lagoon	4-6	83.0	7.4	0.06
3	Lagoon	4-6	81.5	11.9	0.19
4	Lagoon	4-6	86.7	13.6	0.19
5	Lagoon	4-6	83.3	17.5	0.12
6	Lagoon	4-6	88.9	17.2	0.10
7	Densator Underflow	1.6	71.2	35.2	-
8	Densator Underflow	1.5	72.5	21.0	0.34
9	Lagoon		60.2	10.4	0.30
10	Densator Underflow		45.7	42.0	-
11	Densator Underflow	4.5	63.8	98.5	0.17
12	Lagoon		81.4	21.5	0.11
13	Lagoon		80.0	10.4	0.54
14	Lagoon		80.0	5.2	0.24
15	Densator	3.7	81.7	68.3	0.24

Notes:

[a] These solids rates include the weight of the filter aid, which will constitute 10 to 40% of the weight over the range of 6-100 lbs/ft²/24 hrs.

[b] Feed slurries in all tests were mixtures from several types of mine water treated with partially carbonated hydrated lime.

[c] Although the % solids in these feed slurries were high, the filtration rates were poor. This sludge came from compartment two of the settling lagoon where a very fluffy, slow settling component of the original sludge settled.

Table 55. CHEMICAL ANALYSES OF FILTER CAKE FROM CMD SLUDGE (weight % (dried at 105°C))

Test no.	Iron	Aluminum	Calcium	Sodium	Sulfate
1	18.1	5.4	3.2	0.88	3.7
2	20.1	5.9	3.1	0.91	2.1
3	20.4	6.7	3.1	0.82	2.0
4	28.7	4.6	5.0	0.34	4.5
5	21.4	4.7	4.4	0.76	3.9
6	26.2	5.5	5.9	0.24	5.5
7	22.0	4.5	4.3	1.05	2.2
8	9.8	3.6	2.3	1.69	0.7
12	23.9	6.0	2.7	0.51	5.6
13	4.8	3.9	16.9	0.62	1.2
14	17.5	6.3	2.8	0.84	3.1

Precoat Requirements

Plant experience indicated a precoat usage rate of 50 lbs/hr for the 113 ft^2 drum, or 0.4 $lbs/ft^2/hr$. Continuous, uniform operation may reduce this requirement to 0.3 $lbs/ft^2/hr$.

A loss of 100 to 200 pounds of precoat may result from residual suspension in the filter tank after precoating. Provision should permit its recovery in the precoat slurry feed tank for use in the next precoating cycle. Another precoat loss can result at the completion of a precoat loading cycle. When a precoat layer of about 3/4-inch remains, further removal with the cutting knife could result in damage to the filter cloth. It is difficult to add another layer evenly directly on this residual layer. It is preferable to drop this residual layer into the filter tank and provide for its inclusion with the new precoat slurry.

Filter Cake Handling

The filter cake prepared with the plant filter had a moisture content of about 80 percent. This cake fell freely from the cutting knife on to a moving rubber conveyor belt for discharge. No difficulties were experienced with the filter cake sticking to the belt. In subfreezing weather, the filter cake tended to freeze quickly creating handling problems in truck transportation.

Filtration Using Flocculants

The effect of flocculants on the filtration characteristics of several sludges were made in bench scale and plant tests. The filter test leaf data are presented in Table 56. At flocculant levels of 20 to 40 ppm (2 to 3 lbs/ton dry solids), an extensive agglomerating effect caused a very rapid separation of liquid and solids in the treated slurry. Formation of these ball-like flocs should produce much higher filtration rates. There were higher liquid rates but no improvement in solids filtration rates. The low solids filtration rate occurred when large portions of the filter cake fell from the filter media surface as it emerged from the liquid feed bed. In continuous practice with the high liquid filtration rate, the solids in the filter bed would soon reach an unmanageably high level. The flocculants created a slippery, non-cohesive sludge flow. The cohesiveness decreases with higher flocculant dosage (see Tests 5, 6, 12, 13, and 29 in Table 56).

Plant Tests with Strongly Flocculated Feed Slurries

The data from the full-size precoat filter using strongly flocculated pulps are presented in Table 57. The runs at each dosage level were short, and operating difficulties prevented any tests at the highest drum speeds. The results were in agreement with those from the bench scale test leaf. The most significant observations made were:

1. Strongly flocculated filter feeds do not respond as well as weakly flocculated slurries with a conventional precoat filter, even when pulp suspension is maintained.

2. At high flocculant dosages, the solids settle very fast to the bottom of the filter tank, providing little contact of the drum with the solids. In a separate operation, slurries could be dewatered in this manner to attain a heavier and more homogeneous pulp for filter feed slurry.

3. In contrast to the test leaf results, high liquid filtration rates were not attained while using strongly flocculated slurries.

Correlation of Bench Scale and Plant Operations with Design Criteria

In general, the test leaf results could be extrapolated directly to predict performance of the full scale vacuum filter. The major filter design criteria for a low cost vacuum filter operation, which were developed and confirmed in both bench and plant, are summarized in Table 58.

Table 56. SUMMARY OF FILTRATION LEAF TESTS USING FLOCCULANTS

All tests made with an Eimco 0.1 ft^2 test leaf.
Temperature: 70° F
Vacuum: 24 inches Hg

Filter Cycle: 16 sec submergence
50 sec drying

Test No.	Flocculant Used[a] Type	Dosage[b] ppm	Dosage[b] lbs/ton dry solids	Filter Aid Used[e]	Type of Slurry Loading to Filter Leaf	Filtration Rates Solids lbs/ft^2/24 hrs	Filtration Rates Liquid gpm/ft^2
Proctor No. 2 - 2.71% Solids Feed - Dolomitic Pebble Lime							
1	None	(Controls)		Precoat	Normal[c]	53.2	0.15
2	None	(Controls)		Precoat	Normal	53.8	0.15
3	1A1	20	1.47	Precoat	Normal	52.0	0.16
4	1A1	40	2.94	Precoat	Normal	53.2	0.16
5	1A1	80	5.88	Precoat	Normal	33.8	0.27
6	1A1	80	5.88	None	Normal	25.2	0.19
7	1A1	80	5.88	None	Top Load[d]	100	0.30
8	None	(Control)		None	Top Load	34.6	0.12
Proctor No. 1 - 2.53% Solid Feed							
9	None	(Controls)		Precoat	Normal	54.0	0.18
10	None	(Controls)		Precoat	Normal	53.0	0.18
11	1A1	20	1.58	Precoat	Normal	49.5	0.18
12	1A1	40	3.16	Precoat	Normal	40.3	0.23
13	1A1	40	3.16	None	Normal	33.4	0.30
14	1A1	40	3.16	None	Top Load	118.3	0.36
Bennetts Branch - 1.96% Solids Feed - Fresh Lime Hydrate							
15	None	(Control)		Precoat	Normal	33.1	0.16
16	1A1	10	1.02	Precoat	Normal	40.5	0.30
17	1A1	20	2.04	Precoat	Normal	41.7	0.53
18	1A1	10	1.02	None	Top Load	51.1	0.24
19	1A1	20	2.04	None	Top Load	75.9	0.42
20	1A1	20	2.04	Precoat + 1% Body Feed	Normal	68.2	0.42
21	1A1	20	2.04	Precoat + 1% Body Feed	Top Load	94.0	0.42
22	None	-	-	None	Normal	38.6	0.16
Proctor No. 2 - 7.18% Solids - Old Lime Hydrate[f] - Biochemical Oxidation[g]							
23	None	(Control)		Precoat	Normal	331.0	0.33
24	1A1	20	0.56	Precoat	Normal-B.T.[h]	400.0	0.37
25	1A1	20	0.56	Precoat	Normal-B.T.	326.0	0.43
26	1A1	20	0.56	Precoat	Normal-B.T.	290.0	0.33
27	1A1	20	0.56	None	Normal	264.0	0.37
28	1A1	20	0.56	None	Top Load	370.0	0.38
29	1A1	30	0.84	None	Normal	92.5	0.69
30	1A1	30	0.84	None	Top Load	232.0	0.57

Notes:

[a] ATLASEP

[b] Flocculant dosage indicates ppm per slurry volume (e.g., 20 ppm = 20 mg/liter of slurry. To convert to lbs/ton dry solids:

$$\text{lbs flocculant/ton dry solids} = \frac{\text{Dosage (ppm)}}{5 \times \%\text{ solids in slurry}}).$$

[c] "Normal" loading is the conventional cake pick up obtained by dipping the inverted test leaf into the slurry.

[d] "Top Load" is the technique of pouring slurry into the upright filter leaf, simulating a Buchner funnel test.

[e] Celite 545.

[f] This alkali was high calcium hydrated lime containing 50% $CaCO_3$ from CO_2 absorption.

[g] This water was neutralized in the Densator.

[h] The settled slurry in the bottom of the settler is more dense with high flocculant doses. B.T. = bottom of tank; T.T. = top of tank.

Table 57. PILOT PLANT PRECOAT ROTARY VACUUM FILTER TESTS WITH STRONGLY FLOCCULATED FEED SLURRIES

Test No.	Drum Speed		% Solids in Feed	Flocculant Dosage lbs/ton dry feed solids	% Solids in Wet Cake	Filtration Rates		
	Cycles/day	RPH				lbs/ft^2 per hr.	lbs/ft^2 per day	Liquid GPM/ft^2
1	806	33.6	7.53	0.95	20.8	2.5	59.7	0.21
2	806	33.6	3.71	0	18.3	2.8	68.3	0.21
3	806	33.6	3.71	3.03	19.2	3.6	86.1	0.21
4	806	33.6	3.71	9.9	20.2	1.5	36.2[1]	0.23
5	410	17.1	0.84	1.41	19.0	0.6	14.6	0.36

Vacuum - 22" Hg.
Filter Area - 113 ft^2
Filter Aid - Celite 545 (2" cake)
Filter Aid Consumption - 50-60 lbs/hr.
Filtrate Quality - Clear

Flocculant Used - ATLASEP 1-A-1
Sludge Origin - Proctor No. 2 water subject to biochemical iron oxidation, hydrate lime neutralization.

Note:

[a] The flocculant level used in Test 4 settled sludge to the bottom of the filter vat, preventing pickup by the filter drum.

Table 58. RECOMMENDED VACUUM FILTER DESIGN PARAMETERS

1. Optimum solids content of filter feed - 3% solids minimum.
2. Liquid filtration rates - 0.35 gpm/ft^2 - maximum (70°F)
   0.15-0.20 - average (70°F).
3. Solids filtration rates - 15 to 25 lbs/ft^2/24 hrs. @ 1% solids feed.
   120 lbs/ft^2/24 hrs. @ 4% solids feed.
   (all values for 70°F)

   A graph relating the entire range of filtration rates vs. solids content in the filter feed is presented in Figure 38.
4. Filter drum speed - 0.8 rpm minimum.
5. Drum submergence - 25%.
6. Precoat consumption - 40-50 lbs/hr for 113 ft^2 filter area.
   0.4 lbs/hr/ft^2 filter area.
   40% of dry sludge weight @ 24 lbs/ft^2/24 hrs.
   8% of dry sludge weight @ 120 lbs/ft^2/24 hrs.
7. Knife advance - 4 mils/drum revolution.
8. Moisture in filter cake - 76 to 82%.

   Cakes of this moisture content could be handled satisfactorily on a conveyor belt.
9. Specific gravity of filter feed - 1.03 (@ 4% solids).
10. Specific gravity of filter cake - 1.13 (71 lbs/ft^3).
11. Type of filter cloth - EIMCO P.O. - 801 HF or equivalent.
12. Flocculant requirement - 2 to 4 ppm added to Densator feed.
    (will vary with flocculant)
13. Filter aid type - Celite 545 or equivalent.

The recommended design parameters should produce satisfactory, economical filter operation if a precoat filter is to be considered. However, the preliminary test work without filter aid with strongly flocculated pulps suggested more study seeking to eliminate the use of filter aid and its high labor requirements.

In sizing a precoat filter, the capability to store sludge and the frequency and extent of filter operation requirements must be related to sludge volume and filtration rates. Once precoated, a drum filter must be kept under vacuum with continual moisture conditions to maintain the filter media on the drum until it is all consumed. The power requirements for maintaining the vacuum represent a major portion of filtration costs, comparable only to the costs for precoat material.

Filtration Costs

Filtration costs for coal mine drainage slurries are determined by the following factors:

1. The maximum liquid filtration rate attainable, which was found to be 0.3 GPH/ft^2.
2. The maximum solids filtration rate attainable at maximum liquid filtration rate (a function of the solids content of the filter feed), which filter leaf tests indicated could be as high as 300 lbs/ft^2/24 hrs with hydroxide-produced sludges.
3. The solids content of the filter feed, which both pilot plant and bench scale tests indicated, could be as high as 15% with carbonate-produced sludges.
4. The size of the filters required, and their construction materials.
5. The amount of filter aid required, which was established on the pilot plant filter to be about 0.4 lbs/hr/ft^2 filter area.

Using the operating range as established in these evaluations, the costs of filtration at various solid filtration rates were calculated and presented in Table 59. Filtration equipment cost variation with sludge product rates and filter area may be derived from Figures 38 and 39.

The cost estimates of Table 59 indicate that reduction in filtration costs is attainable with higher feed solids in contrast to the so-called "normal" conditions (1% solids feed or lower) reported by others[72, 75, 98]. For cost comparisons with other secondary sludge

Table 59. ESTIMATED FILTRATION COSTS FOR VARIOUS LEVELS OF SOLIDS FILTRATION RATES

Basis: 1,000,000 GPD mine water ccntaining 1,000 ppm neutralized solids (4.15 tons/day of dry sludge). 365 operating days per year.

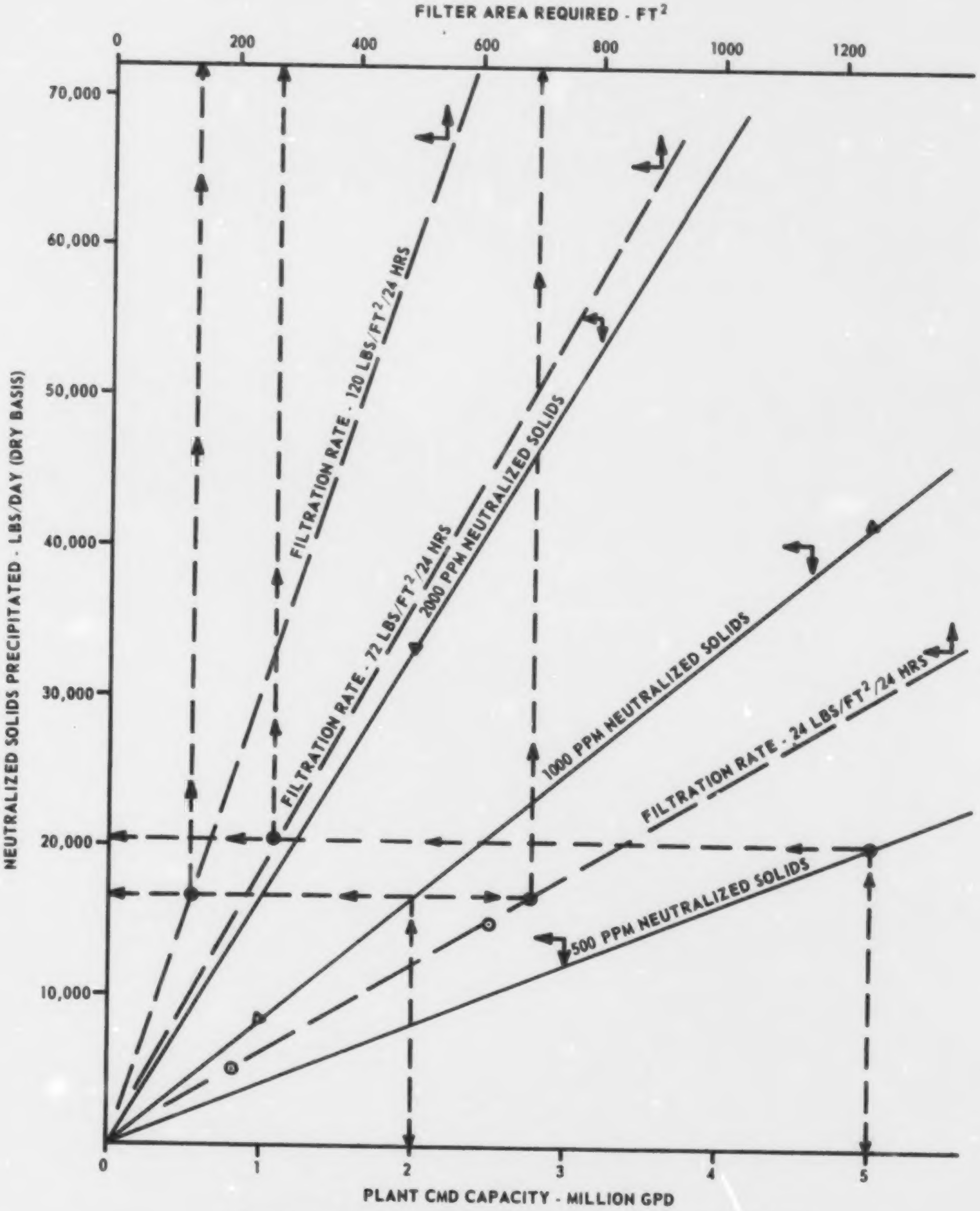
Solids Filtration Rate - Lbs/Ft^2/24 Hrs	A 24	B 120	C 240
1. % Solids in Filter Feed	1	4	8
2. Filter Area Required - Ft^2	340	68	34
3. Installed Filter Cost (mild steel)	$140,000	$35,000	$20,000
4. Annual Amortization (10% per year)	$ 14,000	$ 3,500	$ 2,000
5. Filter Aid Consumption (0.4 Lbs/ft^2/Hr)-Lbs[a]	3,260	650	325
6. Daily Costs			
a. Amortization	$ 38.00	$ 10.00	$ 6.00
b. Electric Power (12¢/Ft^2/Day @ 1.5¢ KWH)	41.00	8.00	4.00
c. Filter Aid (@ $117/ton)[b]	191.00	38.00	19.00
d. Flocculant (@ $1.20/Lb)	---	5.00	10.00
e. Operating Labor (1 man-1/3 time @ $3/hr)	24.00	24.00	24.00
f. Maintenance (5% installed cost/yr)	19.00	5.00	4.00
g. TOTAL OPERATING COST - DAILY	$ 313.00	$ 90.00	$ 67.00
7. Cost per 1,000 gal. CMD treated	31.3¢	9.0¢	6.7¢
8. Cost per ton dry solids filtered	$ 75.50	$ 21.70	$ 16.10
9. Cost of Filter Aid per 1,000 gal. treated	19.1¢	3.8¢	1.9¢
Cost of Filter Aid per ton solids treated	$ 46.00	$ 9.15	$ 4.60

Notes:

[a] This use rate might be decreased to 0.3 Lbs/Ft^2/Hr.

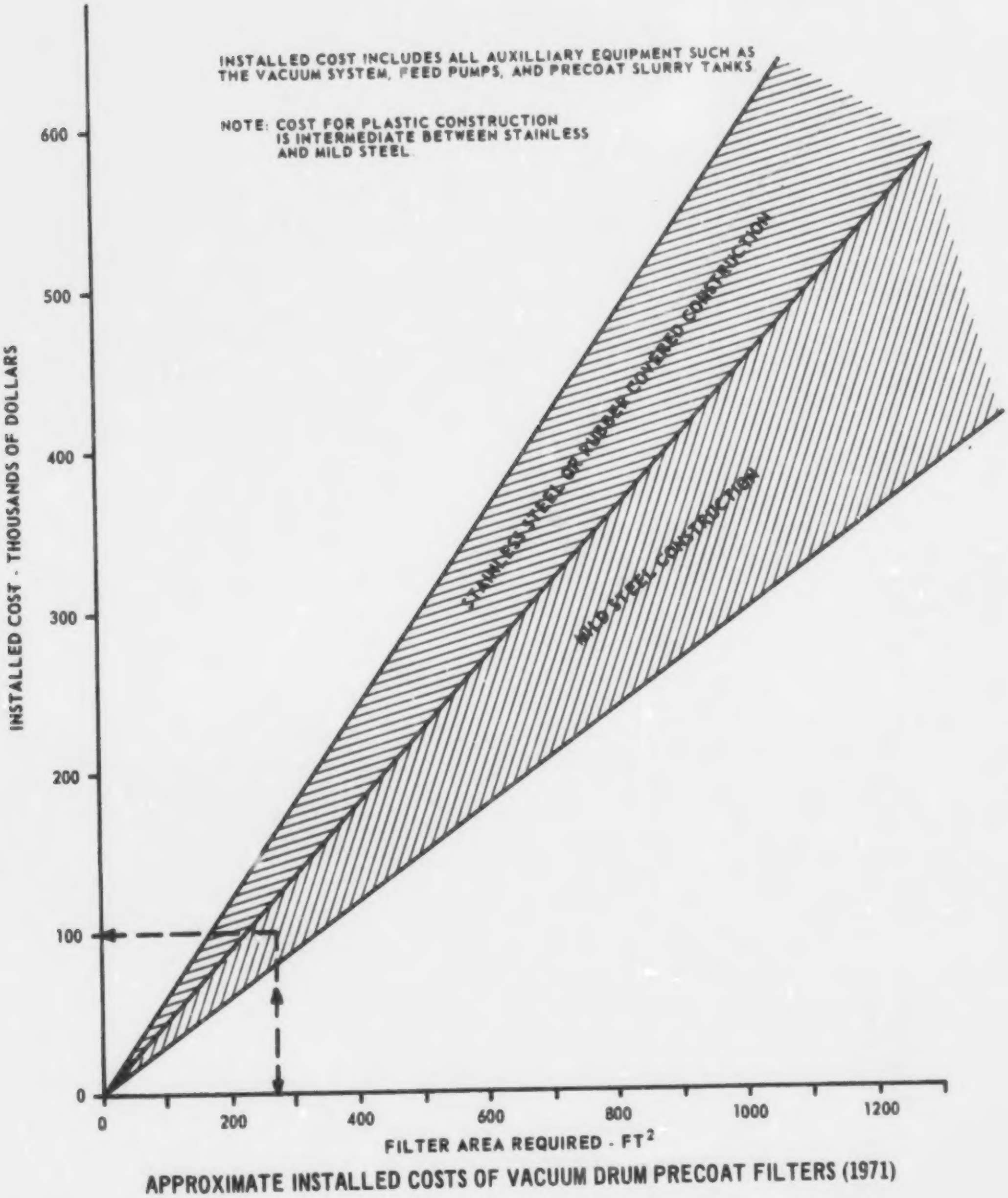
[b] Based upon a minimum carload lot of 60,000 Lbs @ $84.00/Ton F.O.B. California, plus $33.50 freight to Pennsylvania.

Figure 38. Sludge production rates and filter area requirements



SLUDGE PRODUCTION RATES AND FILTER AREA REQUIREMENTS

Figure 39. Approximate installed costs of vacuum drum precoat filters (1971)



APPROXIMATE INSTALLED COSTS OF VACUUM DRUM PRECOAT FILTERS (1971) dewatering methods, the cost per ton of dry solids processed are plotted in Figure 40. Present treatment activity indicates a dry solids production range of 1 to 25 tons/day.

A study of filtration costs permits the following conclusions:

1. Filtration cost for sludges can range from $15/ton of dry solids to $90/ton, using a conventional vacuum drum precoat filter.

2. The conventional precoat filtration cost would probably never be less than $10/ton dry solids filtered.

3. Filter aid is the major cost item, constituting about 50% of the total cost.

4. If, with flocculants, filtration rates of at least 240 lbs/ft^2/24hrs could be attained on top-loading type filters with the elimination of filter aid, filtration costs could probably be reduced to about $7/ton dry solids.

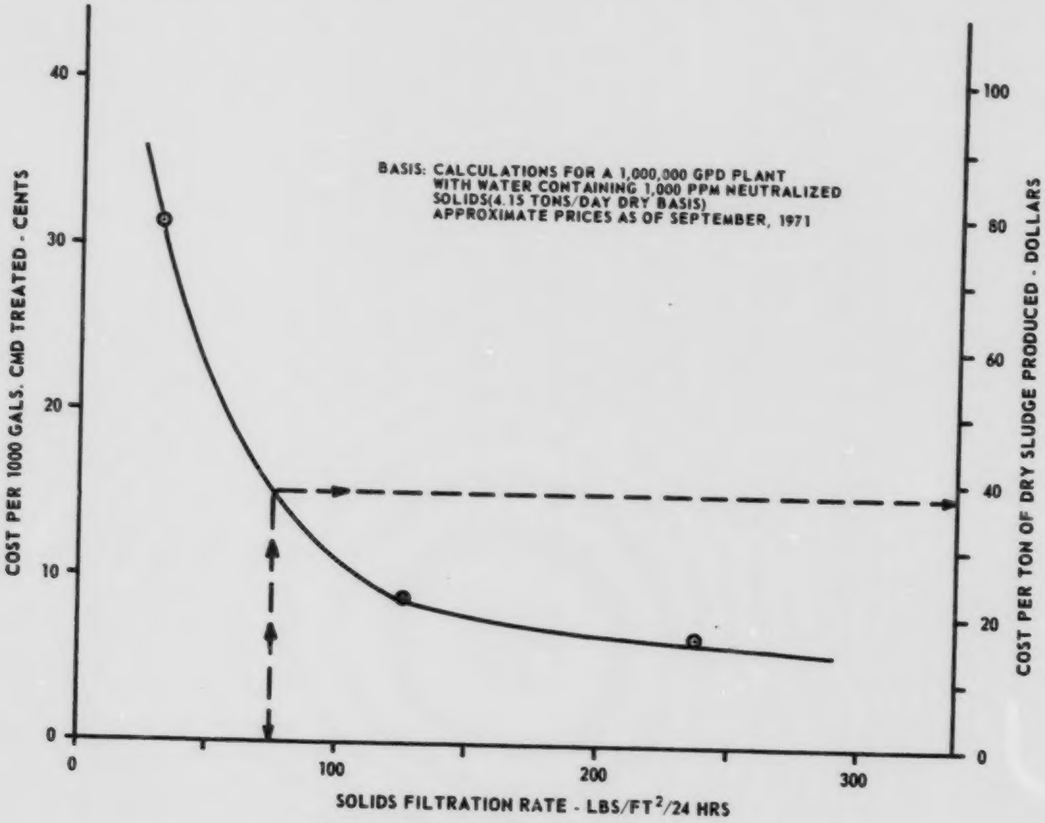
Centrifuge Tests

Performance sludge dewatering experience was attained with a Dorr-Oliver Merco Bowl Centrifuge (Model Z-1-L). It has a maximum bowl diameter of nine inches and an inside length of 24 inches. This is a horizontal, continuous flow machine. The solids settle and compact against the walls under centrifugal force (up to 5200 gravities). They are conveyed from the liquid pool for further drying under centrifugal force. The clarified centrate moves counter current for pump discharge at 30 psi.

The sludge dried (primarily of limestone reaction origin) was pumped from the bottom of the settling lagoon to a storage feed tank and maintained by slow stirring as a uniform feed to the centrifuge. Prior experience[72, 75] assisted selection of operating conditions and maintenance of reasonable clarity in the centrate. The test conditions and results are given in Table 60. These data are in agreement with previous results with the same machine[72, 75]. At higher flows the turbidity of the centrate increased markedly. Should this clarity be a problem the relatively small volume could be recycled through the process thickener or settling lagoon. The centrifuge cake had satisfactory handling characteristics although it showed a tendency to flow and was more subject to a loss of physical integrity than the filter cake.

The device functions satisfactorily and holds great advantage in ease of operation with a minimum of manpower requirements. As a high speed mechanical device, it is subject to high maintenance

Figure 40. Costs of vacuum drum precoat filtration as a function of solids filtration rate capacity



BASIS: CALCULATIONS FOR A 1,000,000 GPD PLANT
WITH WATER CONTAINING 1,000 PPM NEUTRALIZED
SOLIDS(4.15 TONS/DAY DRY BASIS)
APPROXIMATE PRICES AS OF SEPTEMBER, 1971

COSTS OF VACUUM DRUM PRECOAT FILTRATION
AS A FUNCTION OF SOLIDS FILTRATION RATE CAPACITY

Table 60. OPERATING PARAMETERS FOR THE MERCO BOWL CENTRIFUGE

	Run #1	Run #2
% Solids in Feed Slurry	0.9	11.5
Feed Flow Rate - gpm	1.71	1.76
Centrate - gpm	1.71	1.76
Centrate Solids - ppm	57	31
Cake - % Solids	19.3	21.0
Cake Discharge Rate - lbs/min	0.125	0.775
Flocculant (Atlasep 1A1) Dosage Rate - ppm	5	5
Amperes	11	11
Power Consumption -		
kwh/1,000 gpm Centrate	27.2	28.0
kwh/Ton Dry Cake	3,840	562
Centrifuge Speed - rpm	4,620	4,620
Centrifuge Differential - rpm	20	20
Centrifuge Pool Level - inches	5.9	5.9

ANALYSIS OF CENTRIFUGE FEED SOLIDS

Element	Weight %	Compound	Weight %
Al	7.68	$Al(OH)_3$	22.2
Fe	24.10	$Fe(OH)_3$	46.4
Ca	7.46	Gypsum	8.7
Na	0.12	$CaCO_3$	13.6
SO_4	4.86		
		Total	90.9

and power costs. As with other secondary dewatering devices, its response improves with feed solids content.

Sludge Drying Bed

The sludge drying bed is another approach for secondary dewatering settled sludge to improve their handling characteristics and reduce volume. Water is removed from the solids by decantation from the surface of a newly-filled bed, by percolation through the bottom of the bed, and by evaporation from the surface.

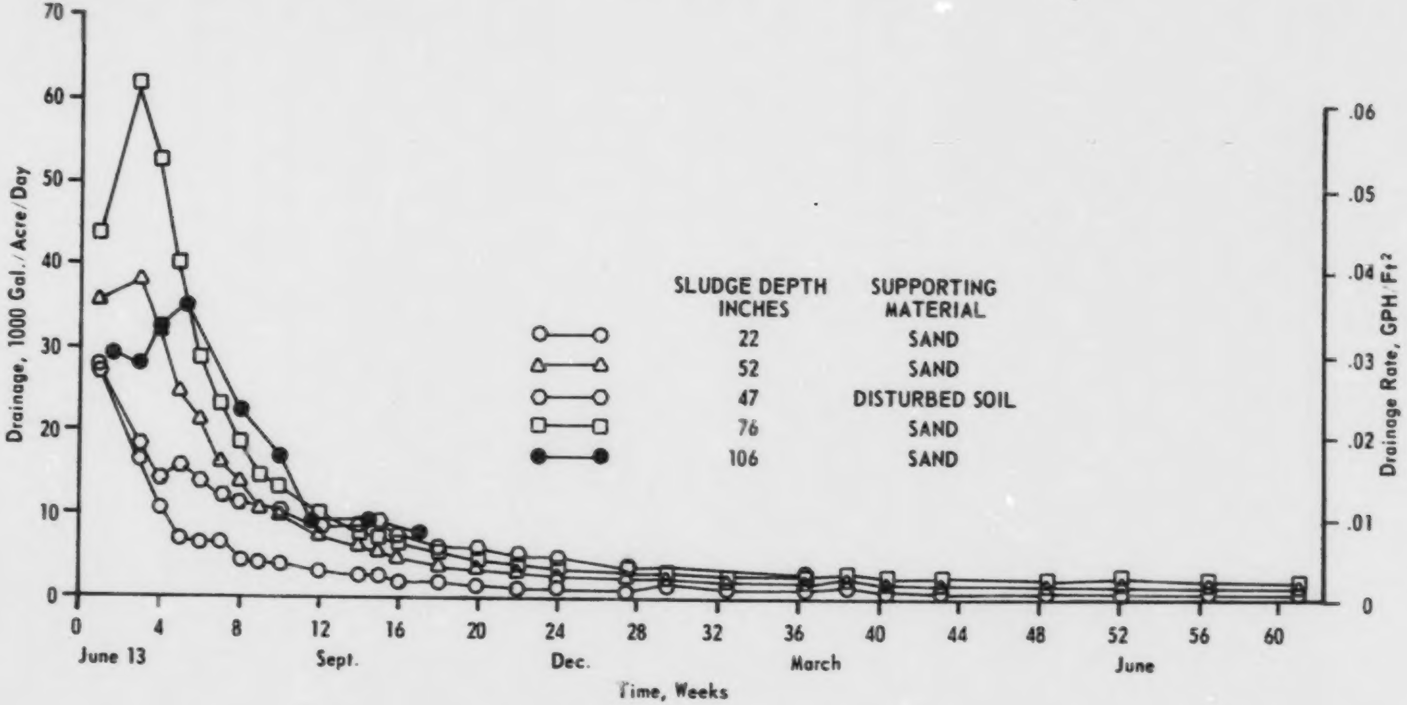
Sludge drying beds have been applied to many types of industrial waste and sewage sludges. It had not been previously considered for CMD sludges. The operating experience reported for sewage sludge dewatering can be related. Of concern is quantitative information regarding mechanisms of dewatering, drainage rates, soil seepage rates, evaporation rates, and regional rainfall.

Jeffrey[83] reported sewage sludge drainage rates and described a procedure for calculating the size of a drainage bed. These drying beds have low solids handling capacity and are normally used on a two year cycle. His drainage rate data are summarized in Figure 41, using an influent sludge slurry with 8 percent solids with several different sand filter bottoms. It was concluded: Initial drainage rates are erratic while the starting layer of sludge is compacting, but the highest rates did not exceed 0.07 GPH/ft^2; The drainage rates stabilized at a level less than 0.01 GPH/ft^2 after 12 weeks' operating time; Stabilized drainage rates were proportional to hydrostatic head; The ultimate practical bed loading rate, when drainage is the only dewatering force, is 2.4 lbs of dry solids per ft.^{2} per 30 days, of drainage; A practical limit to drying by drainage is 75 to 80 percent moisture; Drainage alone produces a sludge mass of uniform moisture content which is independent of sludge depth; The dewatering bed must be located upon a supporting medium with adequate water permeability rates; Evaporation definitely reduces moisture content of the sludge, but the practical effects of evaporation are limited to the top six to 12 inches of sludge; and Since evaporation is necessary to produce a workable sludge, the depth of a sludge bed must be limited to depths at which evaporation can be effective.

These data and actual operation of the drying basin suggest the following factors as to the design and operation of a basin for mine drainage sludge:

1. The decantation stage can be expected to remove much of the water but may require most operator attention to prevent discharge of particulate matter.

Figure 41. Sewage sludge drainage rates



2. The drainage function is essentially a sand filtering operation without backwash capability. The bed capacity is a function of drainage rates through the sludge and the bottom of the bed. This requires a permeable bottom with a controllable water transfer capability.

3. The drainage rate limitation should be that developed by the sludge rather than by the supporting bed. Although drainage rates of supporting soils might be adequate, a specially-constructed sand filter bottom with underlying drainage lines to provide positive, high capacity drainage is preferable. Such a design must consider ground water elevations, direction of drainage flow from the bed, and possible interference from and to natural drainage. Coal refuse banks should be avoided.

4. Evaporation controls the moisture content of the sludge in the final stages. Evaporation rates versus precipitation levels for the bed site are thus critical.

Soil seepage and permeability rate data reported by Todd[84] show drainage rate equivalents ranging from 0.03 to 1.66 gph/ft^2 for sand-clay mixtures to gravelly loam. The rate for coarse sand is 120 gph/ft^2. Figure 42 gives a relationship between decreases in bed depth to bottom drainage rates. The drainage rates for several types of soil which fall within the range of the graph have been indicated.

Data for rainfall and evaporation rates for western Pennsylvania-eastern Ohio are summarized in Figure 43[84] as a 12-month plot. The interaction of these two weather factors is shown as the difference in level of an impervious pond. The graph clearly indicates that appreciable evaporation can be expected during the summer months, but that over the year, they are roughly equivalent. However, in well-drained drying beds the continuous percolation continues during precipitation periods so that rainfall retards only the evaporative process. Obviously a bed cover would minimize these effects and was considered in these studies. Perhaps the more significant weather factor results from sub-freezing conditions which can nearly block drainage. This effect simply requires excess holding capacity for sub-freezing winter operating periods, which seldom exceed two months in this region.

Sludge Drainage Rates - Bench Scale Drainometer Determination--

Only minimal data were available for the design of the bed. The type of information sought for CMD sludge drying bed design and for cost estimation purposes are:

Figure 42. Pond liquid level dropping rate vs. pond bottom drainage rate capacity

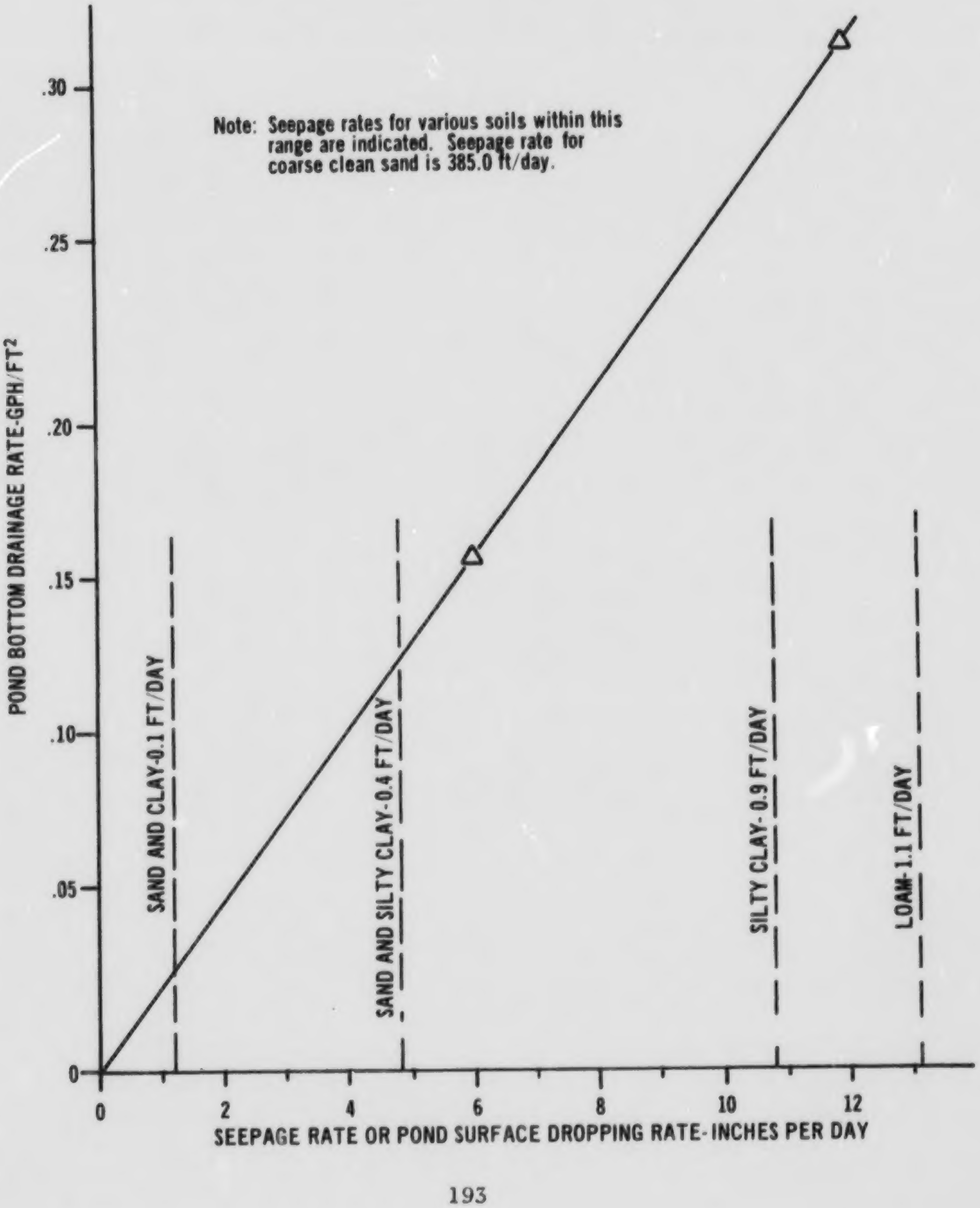
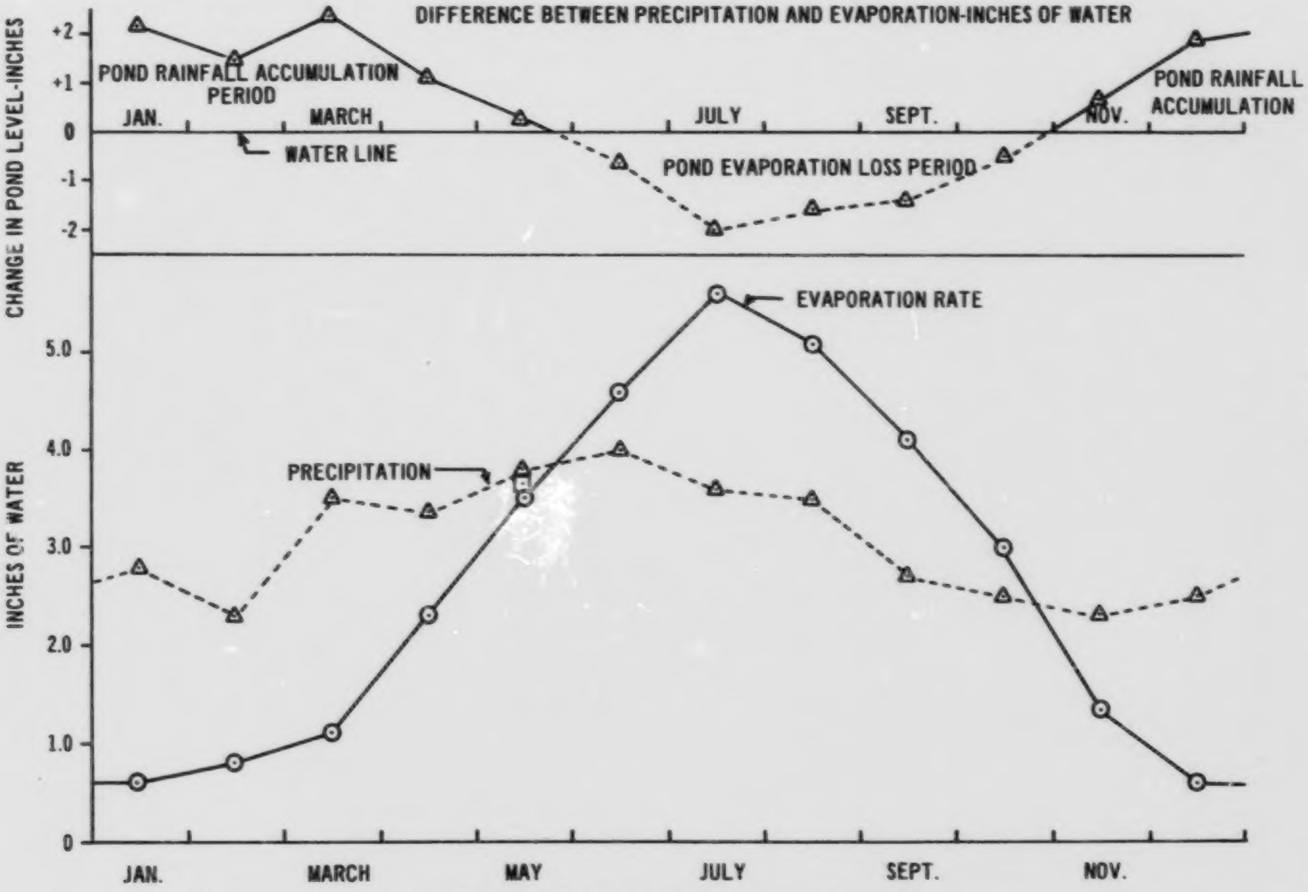


Figure 43. Precipitation and reservoir evaporation data for Western Pennsylvania and Eastern Ohio



BASIS: EVAPORATION RATES-MEAN MONTHLY VALUES
RAINFALL DATA-30-YEAR AVERAGES

1. Drainage rates of various sludges versus time.

2. Permeability rates of water through consolidated sludge.

3. The effect of multiple sludge layers (with intermittent drying) on drainage rates.

4. Effects of sludge surface cracking on drainage rates.

5. Bulk densities and moisture content of the completely drained sludges.

Since the test schedule did not permit production of sufficient sludge from one flowsheet for direct evaluation, a small-scale test unit was developed. A drainometer was constructed from a 100-gallon tank, as detailed in Figure 44.

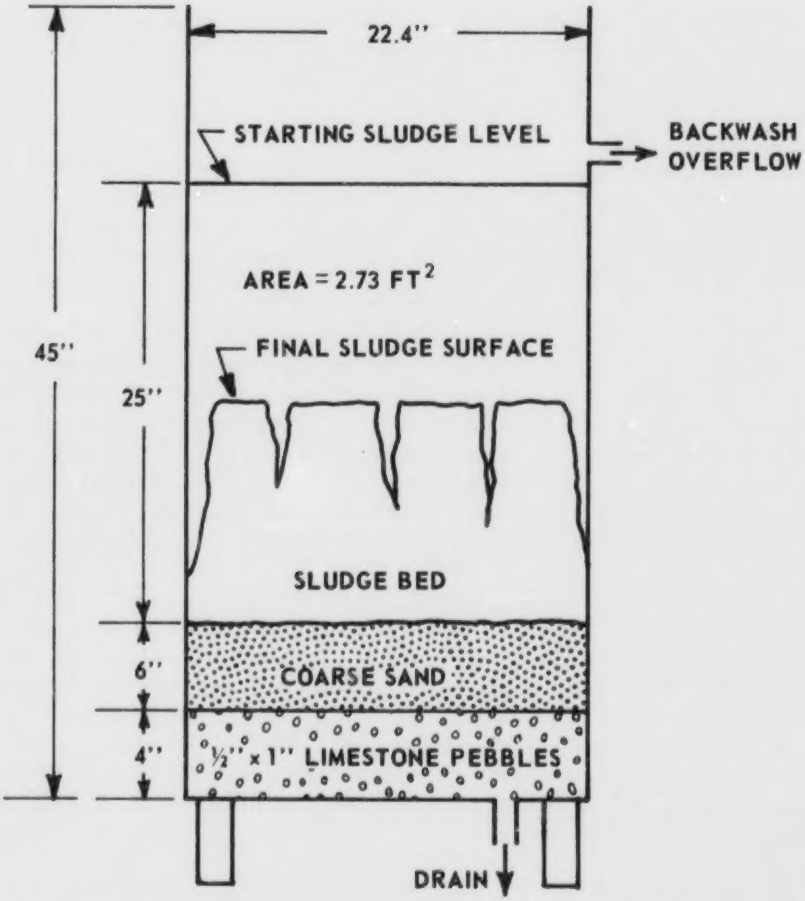
The bottom of the drainometer was filled with 1/2-inch by 1-inch limestone to a depth of four inches upon which was placed a six-inch layer of coarse sand. This simulates an ideal drainage bed medium with the larger stone providing a drainage bed while the sand would trap the sludge particles. Before loading with sludge, the sand bed was backwashed with water to remove any entrapped particles. The clean bed had a drainage rate with water of about two gpm/ft^2 which is comparable for a conventional sand filter. With this design, it was possible to load the drainometer to a sludge depth of about two feet.

In each test, the drainometer was carefully filled with slurry to avoid disturbing the sand bed. The slurry was allowed to drain for a period of about 48 hours, at which time the top of the cake was void of free-standing water and cracks were beginning to form from the decreasing volume of the hydrated particles. The initial drainage rates were measured at hourly intervals and subsequently at longer intervals up to 24 hours total elapsed time. Final rates, which were essentially constant, were measured at eight-hour intervals.

In two tests, a second and third layer of sludge was added to the dewatered layer to simulate the cyclical drying bed loading that was practiced in the full scale bed. Cyclical operations have been cited by the Bureau of Mines[8]. The drainometer seemed to simulate the full scale bed, including the cracking phenomenon on the surface of the dewatered cake. Even the cake, which drained only to six percent solids, cracked as it further air-dried and reduced in volume. The final drained sludges had sufficient integrity to be handled by mechanical bucket.

Each of the sludges tested was produced in the plant equipment as with the filter leaf tests (see the Filtration Section). The results

Figure 44. Drainometer test unit


22.4"
STARTING SLUDGE LEVEL
BACKWASH OVERFLOW
AREA = 2.73 FT²
FINAL SLUDGE SURFACE
45"
25"
SLUDGE BED
6"
COARSE SAND
4"
½" x 1" LIMESTONE PEBBLES
DRAIN

of sludge handling characteristics for filtration and drying bed may be compared for the same sludge. About 100 gallons of settled sludge was withdrawn from the thickener and stored for several days prior to transfer into the drainometer. During this 48- to 96-hour queiscent period under typical conditions experienced in a settling lagoon, none of the sludges (about two feet deep) compressed to a density greater than 2.8 percent solids. This indicated that settling techniques (such as would occur in a settling lagoon) only slowly produce any significant change in terminal sludge density. Chemical analyses of these sludges are presented in Table 50, in the Filtration Section. The drainometer test results are summarized in Table 61. The variations in the drainage rates obtained for lime neutralized sludge from three different waters were limited to the initial part of the cycle and approached nearly identical rates after about 20 hours drainage time. This common rate of about 0.05 gph/ft^2 at one foot hydrostatic head is indicated graphically in Figure 45. The permeability rate of water through 20 inches of 12 percent solids sludge was found to be 0.06 gph/ft^2 at one foot hydrostatic head. The plot of drainage rates for CMD sludges are similar to those reported for sewage sludge (see Figure 41), but were initially more rapid and approached the limiting values more quickly. The drying bed size is influenced by both influent solids content and drainage rates as illustrated in Table 62.

The major conclusions from these tests are:

1. The drainage rate for mine water sludges produced by lime neutralization attains a stable level after 24 hours of about 0.05 gph/ft^2. The drainage rate for limestone treatment sludges is nearly double (0.10 gph/ft^2). Both values are for one foot of hydrostatic head.

2. The water permeability rate of consolidated sludge on a drainage bed bottom is about 0.02 gph/ft^2 at one foot of hydrostatic head.

3. Multiple layers of sludge (with intermittent drying) maintain a drainage rate of 0.03 to 0.05 gph/ft^2.

4. Drainage rates through a cracked dry bed were high initially (about 0.20 gph), but gradually decreased to the 0.05 gph/ft^2 level as the cracks filled with new sludge.

5. Solids content of dewatered sludge after several days of draining ranged from 12-15 percent for hydroxide produced sludges, and 20-30 percent for limestone-neutralized sludges.

6. Shortly after drainage began, the underflow liquids were clear.

Table 61. SUMMARY OF DRAINOMETER TESTS

Temperature: 65°F

Test No.	Raw Water Source	Alkali Used	% Solids in Feed Slurry	Initial Sludge Depth inches	Drainage Rate After 24 hrs gph/ft^2	Drainage Rate After 48 hrs gph/ft^2	% Solids in Final Drained Cake	Bulk Density of Final Sludge lbs/ft^3
1	Proctor No. 1	Hydrated Lime	2.5	25	0.05	0.05	6.0	63.1
2	Tyler	Dolomitic Pebble Lime	1.5	12	0.03	0.02	-	-
3	Proctor No. 2	Dolomitic Pebble Lime	2.7	20	0.05	0.05	11.4	64.7[b]
4	Proctor No. 2	Dolomitic Hydrate	2.7	25	0.05	0.05	15.2 @ Top 21.3 @ Bottom	71.1
5	Proctor No. 2	Hydrated Lime	2.8	20	0.05	0.05	21.2	70.5
6	Proctor No. 2[a]	Limestone	9.3	20	0.10	0.08	30.7	71.2

[a] This water subject to biochemical iron oxidation prior to neutralization in rotary reactor. All other slurries were produced by neutralization in flash mixer and subsequent air oxidation.

[b] Bulk density of bone dry material is 62 lbs/ft^3.

Figure 45. Summary of sludge-drainometer dewatering rates

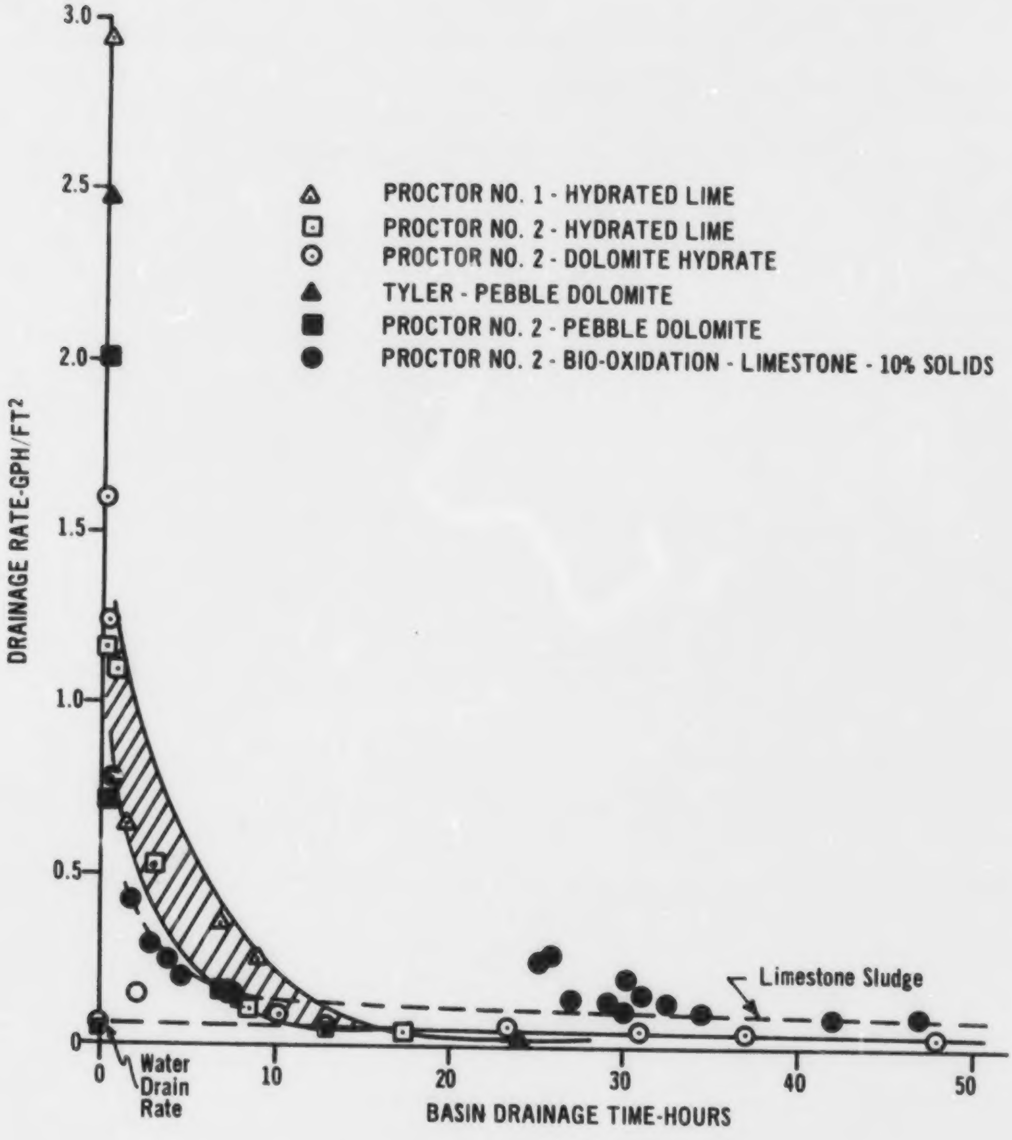


Table 62. EFFECT OF DRAINAGE RATE AND INFLUENT SOLIDS CONTENT ON DRYING BED SIZE

Basis: 820 ft^2 drainage area req'd/1000 GPD bed influent at 2% solids and drainage rate = 0.05 GPH/ft^2

Set No.	Comparative Conditions	% Solids in Influent	Drainage Rate GPH/ft^2	Drainage Area Required ft^2/ 1000 GPD	Decrease in Required Drainage Area	
					ft^2/ 1000 GPD	% of Area
Effect of Increased Influent % Solids at Constant Drainage Rate						
1	A	0.5	0.05	820		
	B	2.0	0.05	720	100	12.2
2	A	2.0	0.05	720		
	B	5.0	0.05	550	170	23.6
3	A	5.0	0.05	550		
	B	8.0	0.05	375	175	31.8
4	A	2.0	0.05	720		
	B	8.0	0.05	375	345	47.9
Effect of Higher Drainage Rates at Constant % Influent Solids						
5	A	0.5	0.05	820		
	B	0.5	0.25	150	670	81.7
6	A	2.0	0.05	720		
	B	2.0	0.25	135	585	81.2
7	A	5.0	0.05	550		
	B	5.0	0.25	107	443	80.5
8	A	8.0	0.05	375		
	B	8.0	0.25	76	299	79.7
Combined Effect of Increased Solids Content and High Drainage Rate						
9	A	0.5	0.05	820		
	B	2.0	0.25	135	685	83.5
10	A	2.0	0.05	720		
	B	5.0	0.25	107	613	86.0
	AB	2.0	0.05	720		
		8.0	0.25	76	644	89.4

7. Penetration of sludge into the sand bed was less than 1/16 inch.

8. The drainometer used could not determine the effect of hydraulic heads greater than 12 inches on drainage rate. However, the sewage sludge data (Figure 41) indicate that drainage rates are directly proportional to hydraulic head, hence faster drainage rates may be attainable by operating a drying bed at a constant operating head of several feet.

The design of the sludge drying basins sought: Capability for decantation from settled sludge; A sand filter bottom for maximum drainage capacity; Elevation to permit drainage from the sub-bed drain lines to suitable discharge; Multiple compartments to provide for continuous operation by allowing drying to proceed in several compartments simultaneously; Access for sludge removal equipment; A bed cover to minimize effect of precipitation; Splash pads under influent lines to prevent erosion; and Provision for a maximum dewatered sludge level of three feet.

Operating Objectives of Sludge Drying Basin--

With design and operating concepts cited, the drying bed utilization sought further evaluation to compare secondary dewatering approaches, including: Realistic design drainage rates; Identification of factors that control drainage rates (e.g., percentage of solids in influent, methods of addition of sludge to the bed, treatment processes, hydraulic head in bed, etc.); Materials handling experience with the dewatered sludge in removing it from the bed (including equipment); Determining bulk densities and water content of the sludges in bed at various stages of dewatering; Determining bed drying times; Effects of weather on drying bed performance; and Operating and maintenance labor requirements for bed operation.

Pilot Plant Parameters--

The range of operating variables are indicated in Table 63. The test program permitted only limited production of several types of sludges which were introduced into the drying basin. The laboratory drainometer data provided a comparison of response for different sludge types (Table 61).

Sludge Drying Bed Operational Observations--

The sludge was introduced into a single compartment of the bed by gravity flow or by pump, as shown in Figure 48, from the settling lagoon or the thickener. Most sludges tended to settle rapidly when contacting the bottom with the supernatant water selectively moving faster over the previously dewatered sludge. This clear water percolated rapidly to the under bed and drain providing a very high

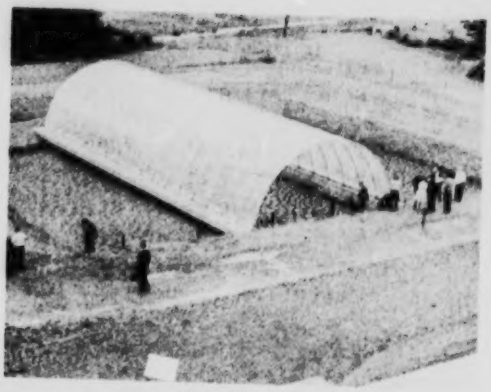


Figure 46. Sludge Drying Basin

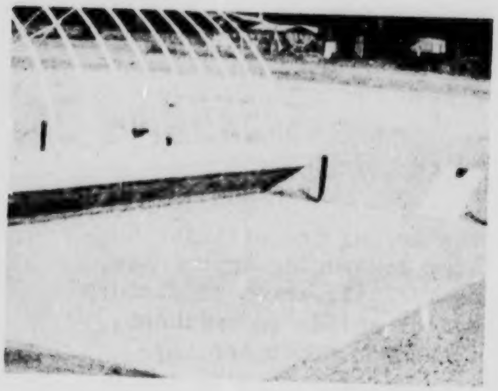


Figure 47. Newly Constructed Sludge Drying Basin Showing Sludge Effluent

Figure 48. Introduction of Settled Sludge into Drying Basin

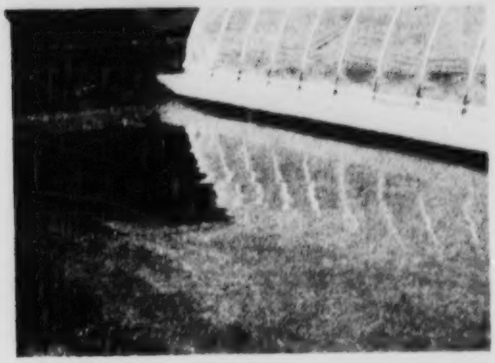


Figure 49. Sludge in Dewatering Process

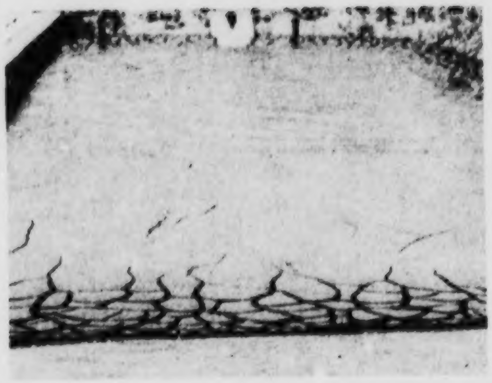


Figure 50. Dewatered-Drying Sludge

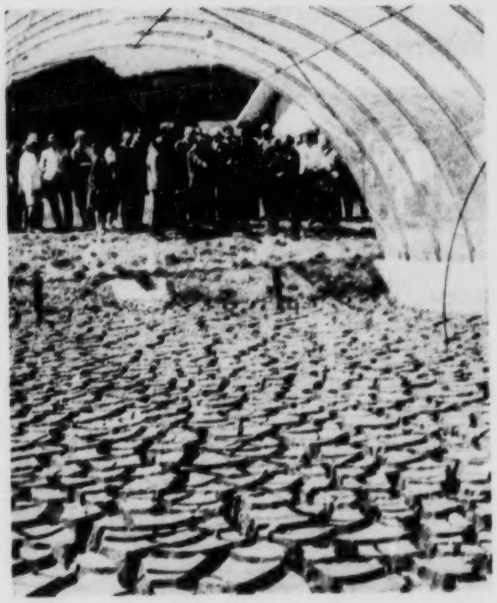


Figure 51. Dried Sludge in Covered Drying Basin

Table 63. OPERATING PARAMETERS LIMITS OF DRYING BED

		Maximum range	Average working range
1.	Slurry working volume - ft^3/section	4,700	---
	Slurry working volume - gallons/section	35,000	---
	Drainage area/section - ft^2	---	1,270
2.	Slurry feed rates - GPM	0-127	25-70
3.	% solids in feed slurry	2-15	2-10
4.	Drainage rates - GPM/section	1-50	1-4
	GPH/ft^2/bottom area	0.03-0.05	0.03-0.06
5.	% solids in dewatered sludge	11-100	13-15
6.	Drainage time for 3 ft depth of feed slurry - days	3-10	5-7
7.	Working sludge depth - ft	0-3	1-2
8.	Weather protection - center section provided with plastic canopy.		
9.	Sludge removal facilities - access for front end loader driven onto drainage bed surface.		
10.	Sludge removal rates (with front end loader) - 2 foot depth of sludge in one section removed in three hours. Equipment: 1 front end loader - 2 cu. yd. 2 trucks - 6 ton and 9 ton 3 men as equipment operators		
11.	Anti-freeze provisions - none.		
12.	Dried solids disposal system - truck haulage to adjacent landfill area.		

initial dewatering rate. This rate varied with the depth and moisture content of the previous sludge layer. A dried sludge layer yielded high initial drainage rates. During the evaporative months, the sludge dried to less than 10 percent moisture (see Figure 51). Under these conditions, the gel-like sludge structure had been broken and rapid percolation followed. This type response favored a slow continuous introduction of the sludge, with longer drying periods.

The optimum sludge depth to be transferred in a given interval was difficult to quantify:

1. A transfer of sludge to the drying bed creating a fresh depth greater than six inches leads to conditions favorable for decantation (the most rapid and effective dewatering procedure) and produces a greater hydraulic head which is conducive to higher percolation rates.

2. The longer interval possible for one filling cycle in one compartment permits longer percolation and evaporative intervals for the other compartments.

3. Evaporation of water from the sludge decreases rapidly with sludge depth after the first six inches. Thus, excessive sludge depths tend to minimize and limit final sludge moisture content.

4. Dry, non-gelled sludge permits rapid percolation rates.

The following qualitative conclusions are indicated regarding the optimum sludge depth:

1. The desirable depth of sludge in the drying bed is a reciprocal function of influent sludge density; i.e. A low density feed sludge may be allowed to accumulate to greater depths since the potential for decantation is greater, the effect of hydraulic head on the drainage rate greater, and yet the final sludge depth will be the least.

2. The dewatered sludge depth desirable is proportional to the seasonal evaporative level; i.e. during periods of high temperature, low humidity, and high wind velocities, the sludge depth may be increased.

3. Some minimum moisture content of dewatered sludge should be attained before a second layer of sludge is introduced.

Decantation of water in the drying basin was practiced to a limited degree by attaching a hose to a board floated on the water surface.

By initiating a siphon flow, the water was transferred to the underdrain air vent. A small pump, also floated, would increase the transfer rate. The supporting board served as an automatic level limitation to prevent the discharge of sludge. Under any conditions, constant manual attention was necessary.

Percolation through the bottom of the bed proceeded rapidly during sludge introduction and slowed as the settled sludge became drainage rate controlling. After completing manual decantation, the drainage continued with a decrease in volume and bed depth (as illustrated in Figure 49) until no free standing water remained. This time interval seldom exceeded 48 hours regardless of sludge depth introduced. Drainage continued at these slower rates, and surface evaporation began. The effect of evaporation first became evident in the shallow sludge layers at the periphery of the bed. The continual decrease in sludge volume increased the growth of shrinkage cracks (see Figure 50) and continued as illustrated in Figure 51 with about 20-inches of dewatered sludge containing 15-30 percent solids. Further decrease in moisture content proceeded very slowly and the effect of sludge depth upon drying became obvious.

It was during the dewatering stages illustrated by Figures 49 and 50 that the plastic bed cover was most effective. After the sludge had dried to the level shown in Figure 51, the cover gave little advantage. Physically, the cover was quite stable and no special problems were experienced. The thin plastic is subject to loss of strength from ultraviolet radiation during the spring and summer months. The weakness develops at any crease or fold and extends upon pressures as from high winds as shown in Figure 4. Replacement appears necessary at six-month intervals. Higher temperatures developed under the canopy were evidenced year around.

The excellent ability of the sand-surfaced bed to retain sludge was evidenced by a very clear underflow and by direct observations of the bed. Figure 53 is a photograph of the bed containing two to three inches of sludge which had reached nearly a bone dry condition. A portion of the sludge and sand was removed. It appeared that the sludge had penetrated less than 1/4 inch into the sand. Unfortunately, upon each removal of the dewatered sludge by Capillar Model 922 High Lift, the sand was disturbed and the underflow developed some turbidity. This had been anticipated and could be readily avoided by replacing the sand. This operation was not attempted since the sludge removal equipment available was somewhat large and difficult to manipulate in these relatively small beds.

When the dewatered sludge reached depths of 24 to 36 inches having at least 15 percent solids, they resisted flow and could be hauled by dump truck without leakage. Rubber-tired equipment did not have sufficient traction to manipulate in the sludge bed, unless the bed was allowed to achieve much higher solids content. Equipment with treads performed well although the 15° access slope seemed excessive when covered with sludge. The use of heavy equipment further disrupted the bed.

The weight of the sludge removal equipment demonstrated an inadequacy of low-cost red dog as a porous bed support. This material did not have sufficient physical stability and was compressed during sludge removal reducing bed drainage rates. The compressed material did protect the clay drain tiles from breakage. Beds probably should utilize physically-stable gravel (such as grade 3A) for the bottom 12 inches followed by a second foot of one-inch stone to support the sand filter layer.

Quantitative Drying Basin Data--

With the basin containing about six inches of dewatered sludge, on August 20, 1971, a sludge settled from limestone treatment of Proctor No. 2 waters was drained by gravity from compartment No. 1 of the settling lagoon to drying basin No. 2. A composite sample contained 8.8 percent solids by weight. Over 13,000 gallons of sludge were transferred in four hours raising the bed depth about 15 inches. This represented a rapid transfer rate, which exceeded sludge production levels. In 20 hours, the bed had dropped nearly 13 inches, more than 0.6 in/hr (about 0.4 g/hr/ft^2). After draining and evaporating ten days, an additional transfer of similar sludge was made to the same compartment which was allowed to dewater for 72 hours when a third transfer of similar sludge was made. The dewatering response of the drying basin under these conditions is detailed in Table 64 and Figure 58. The residual volume and solids content of the dewatered sludge in the drying basin was quite different from that indicated by a sludge settling test (see Figure 59 for the sludge settling data). In the drying basin the residual volume was 24.6 percent, in contrast to the 63.1 percent shown by the settling test. The drainometer tests previsously described provides a satisfactory laboratory approach to the acquisition of such bed drainage data, although giving conservative data.

In further drying operations, following five days of dewatering as described in Table 64, 10,005 gallons of similar sludge (10.6 percent solids) were transferred to drying bed compartment No. 2 which showed a drainage rate of 0.10 g/hr/ft^2 in the next 40 hours and a volume reduction of 65 percent in the next four days. Nearly simultaneously with the above dewatering operation, about 14,000 gallons of the same sludge were transferred to the adjacent drying bed No. 3 which gave a drainage rate of 0.15 gph/ft^2 in the first 40 hours.

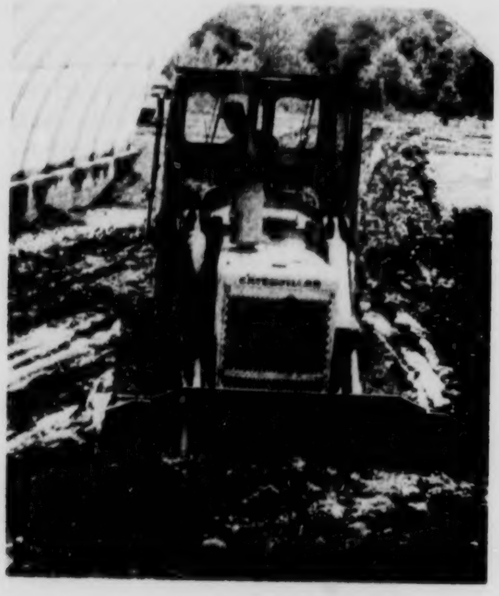


Figure 52. Removing Dewatered Sludge from Drying Basin

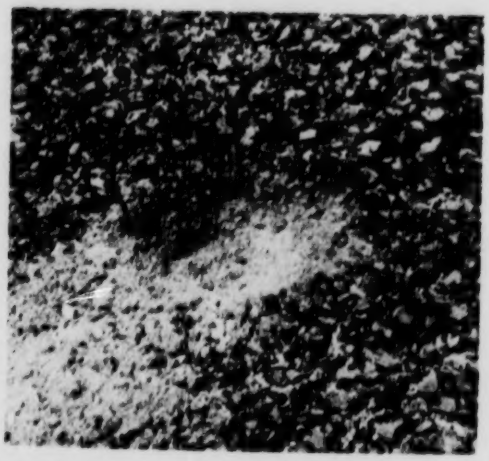


Figure 53. Bone-Dry Sludge Showing Minimal Penetration of Sand

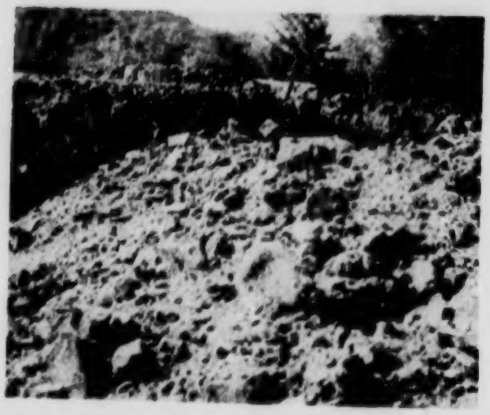


Figure 54. Dewatered Sludge Drying in Land Disposal Area

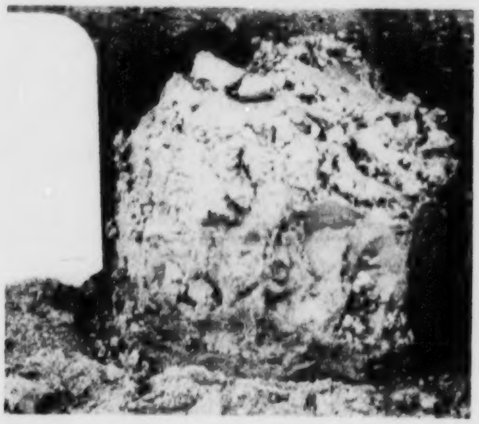


Figure 55. Cubic Foot of Dewatered Sludge

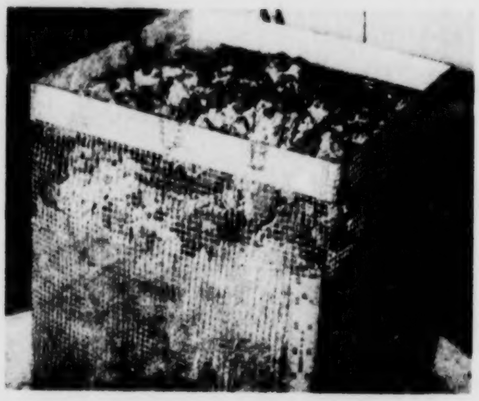


Figure 56. Frozen Dewatered Sludge

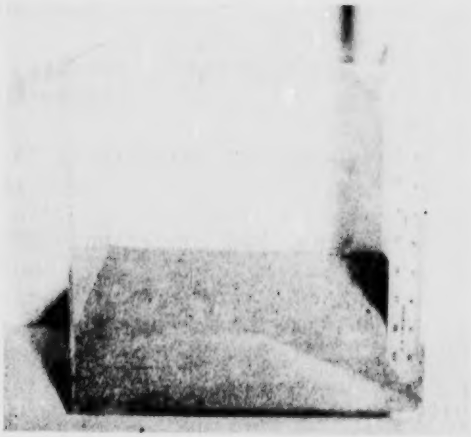


Figure 57. Dried Sludge After Freezing

Table 64. SLUDGE DRYING BASIN EVALUATION

Initial bed conditions:	Drying Basin No. 2 - Covered Depth of Sludge in Bed - Approximately 12 in. Solids Content (Surface) - 15.8% by wt. Drainage period since previous transfer - 72 hours Weather - Temperature - 59-84°F Overcast, humid with rain on fourth day of dewatering
Sludge transfer:	Origin - sludge from settling lagoon, developed from limestone neutralization of Proctor No. 2 water after biochemical oxidation. Pumped from settling lagoon compartment no. 1 to sludge drying basin no. 2. Suspended Solids - 9.5% by wt. (composite) Transfer Rate (average) - 73 gpm for 170 minutes Total Volume Sludge - 12,396 gallons. 11,775 gal. water, 10,315 lb dry solids (using estimated density - 1.05) Bed Level Increase - 14.25 inches
Drainage rate:	Averaged 0.19 g/hr/ft^2 over first 40 hours (see Figure 58 for details)
Sludge drying period:	5 days (121 hours), 97% of total volume reduction in 40 hours
Final bed condition:	Sludge Bed Depth Increase - 3.50 inches Solids Content (surface) - 44.8% by wt. Final Volume Sludge Added - 3,045 gallons (407 ft^3) (Using estimated density - 1.10) - 1,850 gal water, 12,522 lb dry solids
Water rejection:	84%
Operating time:	1 man hour during initial sludge transfer

Table 64 (continued). SLUDGE DRYING BASIN EVALUATION

Sludge dry bed capacity[a]: (3 beds - 36 inches of sludge)	80 days 94,000 gallons dewatered sludge 176 tons dry sludge - 390 tons dewatered sludge 12,500 ft^3 = 463 yd^3 587,000 gallons settled sludge 59,000,000 gallons coal drainage (47% excess capacity of plant design)
Cost to remove dewatered sludge:	\$231.50 (\$0.50/yd^3) 0.31 cents/1,000 gal.

[a] Based on addition of 4 inches dewatered sludge to alternate beds every 3 days, thus a nine day drying cycle. Proctor No. 2 with limestone, 1% settled sludge volume, 350 gpm.

[b] Transfer to land fill - 1/4 mile.

Response of Frozen Sludge--

During periods of sub-freezing temperature, the sludge in the drying basin and the dewatered sludge placed for disposal as land fill will freeze and retard dewatering. As the water in the sludge changes physical state, the sludge structure is modified which leads upon thawing to major reductions in moisture content and volume. Although freezing of the dewatering sludge requires planning in basin design and operation to provide adequate storage volume, the overall response to sludge freezing is favorable. To quantify this response, a sludge freeze-thaw test was made.

A cubic foot of dewatered sludge was cut from the drying basin. The physical instability of the sample led to a bulk volume of 2.05 ft^3 upon transfer to a wire mesh freezing basket (Figures 55 and 56). The latter figure shows the sample after being maintained at 0°C for 118 hours. The sample was allowed to thaw at about 20°C, while observing the changes in weight and water drained from the sample. The resulting data are plotted in Figure 60. The water ceased to drain from the sludge after 143 hours when the original one ft^3 volume had reduced to 0.73 ft^3 and the material was transferred to a tray to prevent sludge loss. At that time the total weight loss was 54.4 percent of which 52.2 percent was from drainage water, and the balance from evaporation. After an additional 28 days drying, the moisture content was less than six percent and

Figure 58. Sludge volume and drainage rates sludge drying bed

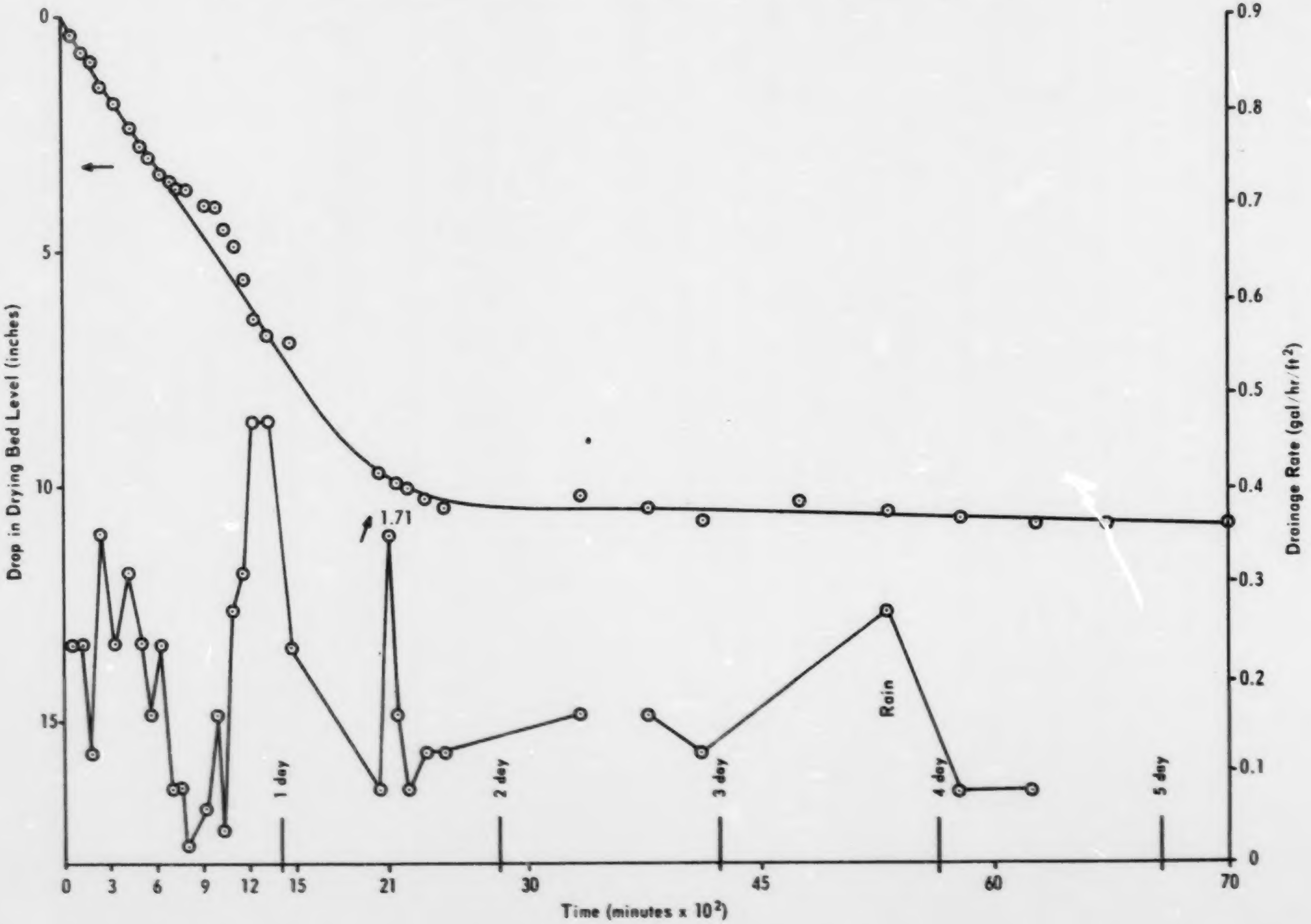


Figure 59. Settling rate of limestone-produced sludge removed from bottom of settling lagoon

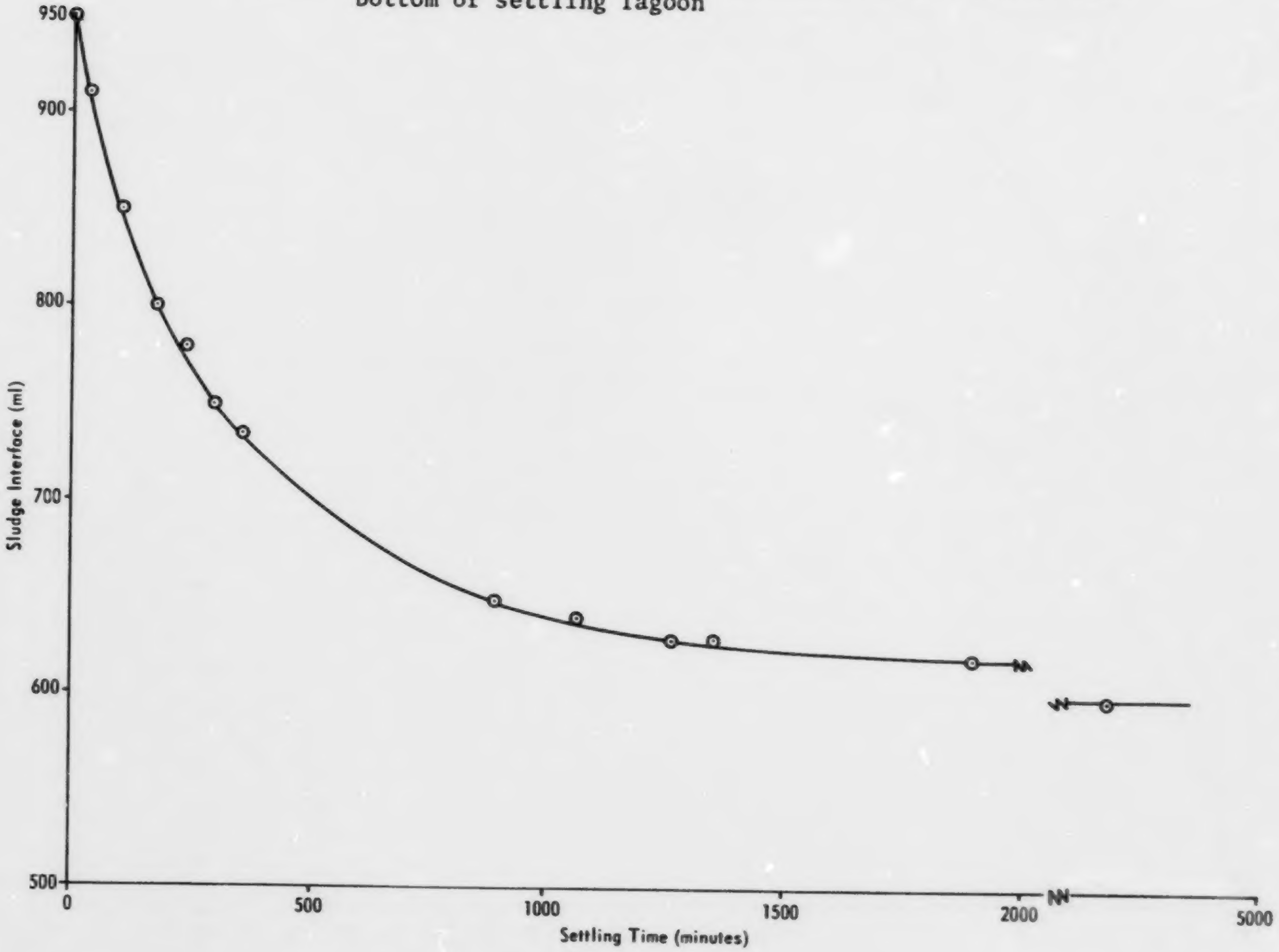
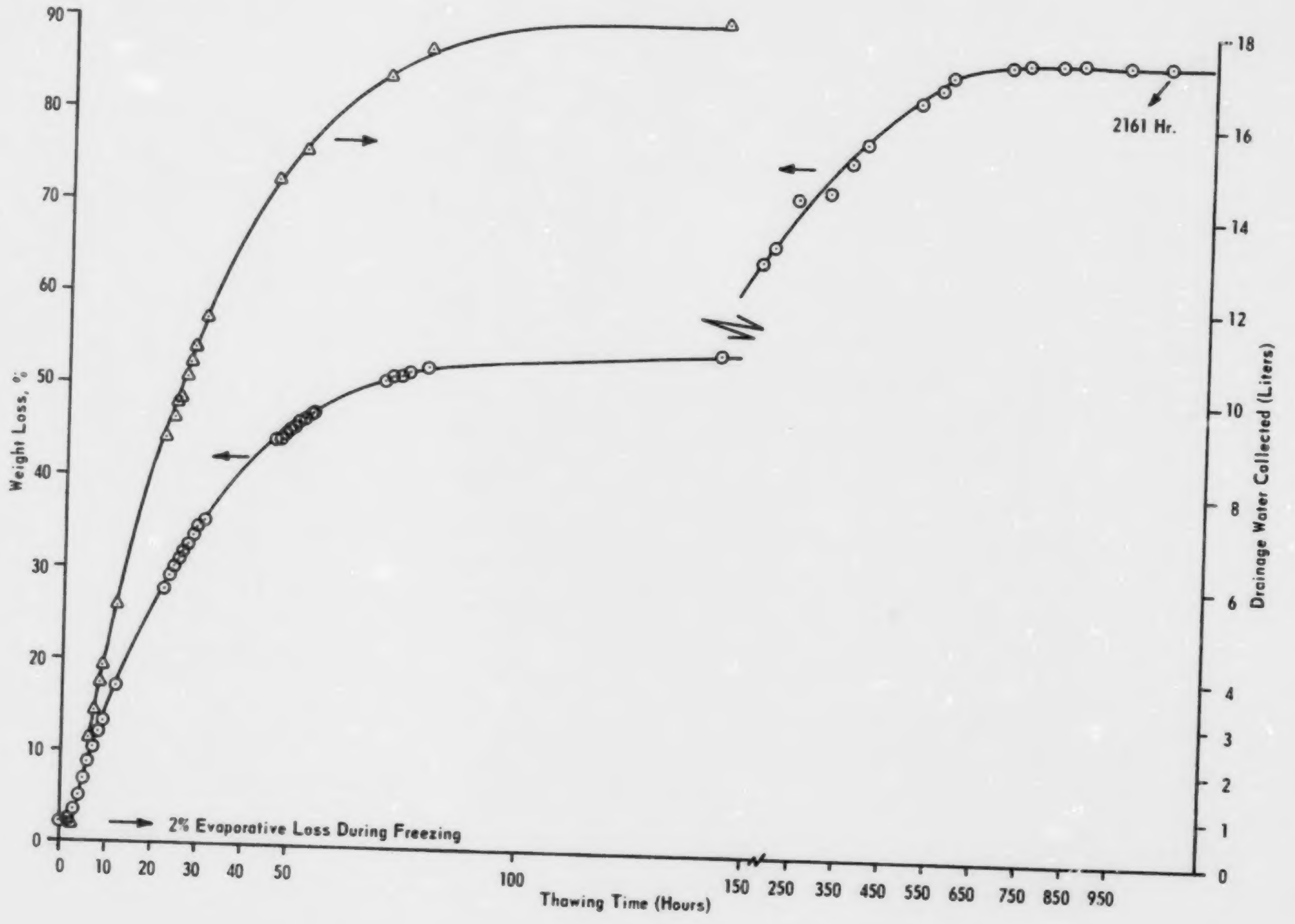


Figure 60. Dewatering rates from frozen sludge



the final volume was 0.25 ft^3 or one-fourth the original dewatered sludge volume (Figure 57). Freezing assists in final dewatering and drying.

Sludge Drying Bed Design Criteria--

The data obtained from drainometer tests and actual drainage bed operation recommended design parameters presented in Table 65. Design graphs for determining bed sizes are contained in Figures 61, 62 and 63.

The drying bed size and number of beds must be established with a selected operating cycle time. Typical bed sizing calculations with their corresponding cycle times for a three foot accumulation of 15 percent sludge are shown in Table 66. Bed size selection must also be dictated by the requirements of the sludge removal system. It appears bed lengths should be limited to 300 feet and widths to about 100 feet. For large bed area requirements use multiple ponds.

Sludge Removal Facilities--

Since sludge removal requires most labor, efficient and low-cost methods of sludge handling should be employed. The sludge must be removed by a system that will not unduly disrupt the sand filter bottom. An option may be the use of a special dragline. The bed size and location should be compatible with a particular dragline configuration.

Most plants would require at least two drying beds for continuous operation. In order for two or more beds to be serviced by a single dragline, a radial pie-segment arrangement is suggested[85]. A potential layout is shown in Figure 64. This system provides for the servicing of two drying beds by a dragline tower located at the center of the quadrant sector and the stacking of the dried sludge around the periphery of the sludge bed area. This arrangement should permit rapid, low-cost removal of the sludge, with dragline power equipment provided at intervals as needed.

The underbed drainage lines are located for servicing using a common sump for control discharge.

Disposal of Dewatered Sludge--

The dewatered sludge attained adequate consistency for truck transportation without leakage, whether dewatered by the drying basin, filter or centrifuge. In all cases only the minimum volume and weight had to be transported. The relative efficiency of handling is better when removing sludge from the drying beds since the truck was used constantly only for the day or two required. In contrast,

Table 65. SLUDGE DRYING BED DESIGN PARAMETERS

1.	Bed Drainage Rate Capacity[a]	0.5 GPH/ft^2 minimum
2.	Percentage of Solids in Feed Slurry	High as possible
3.	Percentage of Solids in Dewatered Sludge	15-30%
4.	Final Sludge Depth[d]	3 ft
5.	Bone-Dry-Solids Capacity	
	(a) for hydroxide slurries	8-12 lbs/ft^3
	(b) for limestone slurries	13-20 lbs/ft^3
6.	Specific Gravity, Feed Slurry	1.03-1.04
7.	Specific Gravity, Dewatered Sludge	1.07-1.13
8.	Bulk Density, Dewatered Sludge	67 to 71 lbs/ft^3
9.	Bed Drying Time(after influent cut-off)	10 days minimum
10.	Total Bed Depth[b]	6 ft minimum
11.	Bed Holding Capacity	30 day production min.
12.	Drainage Area Required	see Figures 61,62,63
13.	Average Bed Drainage Rate Through Sludge	0.03 to 0.05 GPH/ft^2
14.	Number of Beds Required	2 - minimum
15.	Influent Feed Distribution	Even distributed over bed area.
16.	Bed Length	300 ft maximum
17.	Bed Width	See Note c
18.	Evaporation Losses[e]	Varies with weather

Notes:

[a] The bed should contain at least 4 inches of clean coarse sand laid on 8 to 12 inches of 1/2 by 1-inch stone supported by 12-inches 1/4 by 3-inch stone with at least two perforated under-drain tiles running the length of the bed. The 12-foot spacing utilized at Hollywood was adequate.

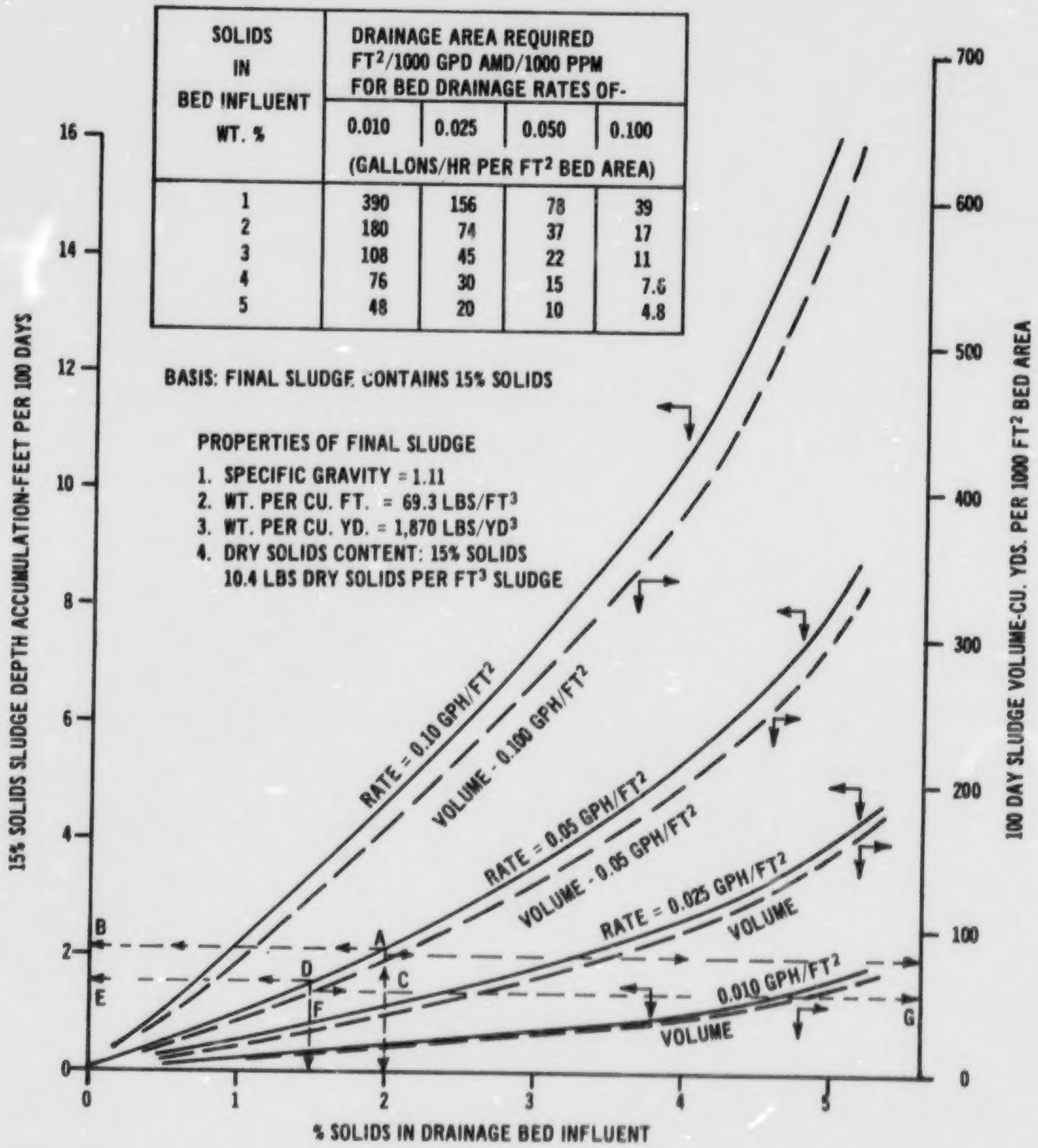
[b] In freezing climates, extra bed depth should be available to provide sufficient bed volume to store 2 to 4 months sludge production during the winter months when drainage rates may decrease drastically.

[c] Bed width should be chosen to be compatible with the type of sludge removal system to be used. A dragline appears desirable for a large bed, hence the bed width should not exceed twice the casting radius of the dragline.

[d] Greater sludge depths may be used, but "drying" times may be excessive for the sludge to attain handling consistency for belt or truck haulage.

[e] Evaporation will aid the final drying of the top 6 to 12 inches of sludge.

Figure 61. Calculator for determining drying basin loading rates and sludge volumes



SOLIDS IN BED INFLUENT WT. %	DRAINAGE AREA REQUIRED FT^2/1000 GPD AMD/1000 PPM FOR BED DRAINAGE RATES OF-			
	0.010	0.025	0.050	0.100
	(GALLONS/HR PER FT^2 BED AREA)			
1	390	156	78	39
2	180	74	37	17
3	108	45	22	11
4	76	30	15	7.6
5	48	20	10	4.8

BASIS: FINAL SLUDGE CONTAINS 15% SOLIDS

PROPERTIES OF FINAL SLUDGE

1. SPECIFIC GRAVITY = 1.11
2. WT. PER CU. FT. = 69.3 LBS/FT^3
3. WT. PER CU. YD. = 1,870 LBS/YD^3
4. DRY SOLIDS CONTENT: 15% SOLIDS
   10.4 LBS DRY SOLIDS PER FT^3 SLUDGE

15% SOLIDS SLUDGE DEPTH ACCUMULATION-FEET PER 100 DAYS
0
2
4
6
8
10
12
14
16
100 DAY SLUDGE VOLUME-CU. YDS. PER 1000 FT² BED AREA
0
100
200
300
400
500
600
700
% SOLIDS IN DRAINAGE BED INFLUENT
0
1
2
3
4
5
RATE = 0.10 GPH/FT²
VOLUME - 0.100 GPH/FT²
RATE = 0.05 GPH/FT²
VOLUME - 0.05 GPH/FT²
RATE = 0.025 GPH/FT²
VOLUME
0.010 GPH/FT²
VOLUME
A
B
C
D
E
F
G

Figure 62. Drainage basin area calculator

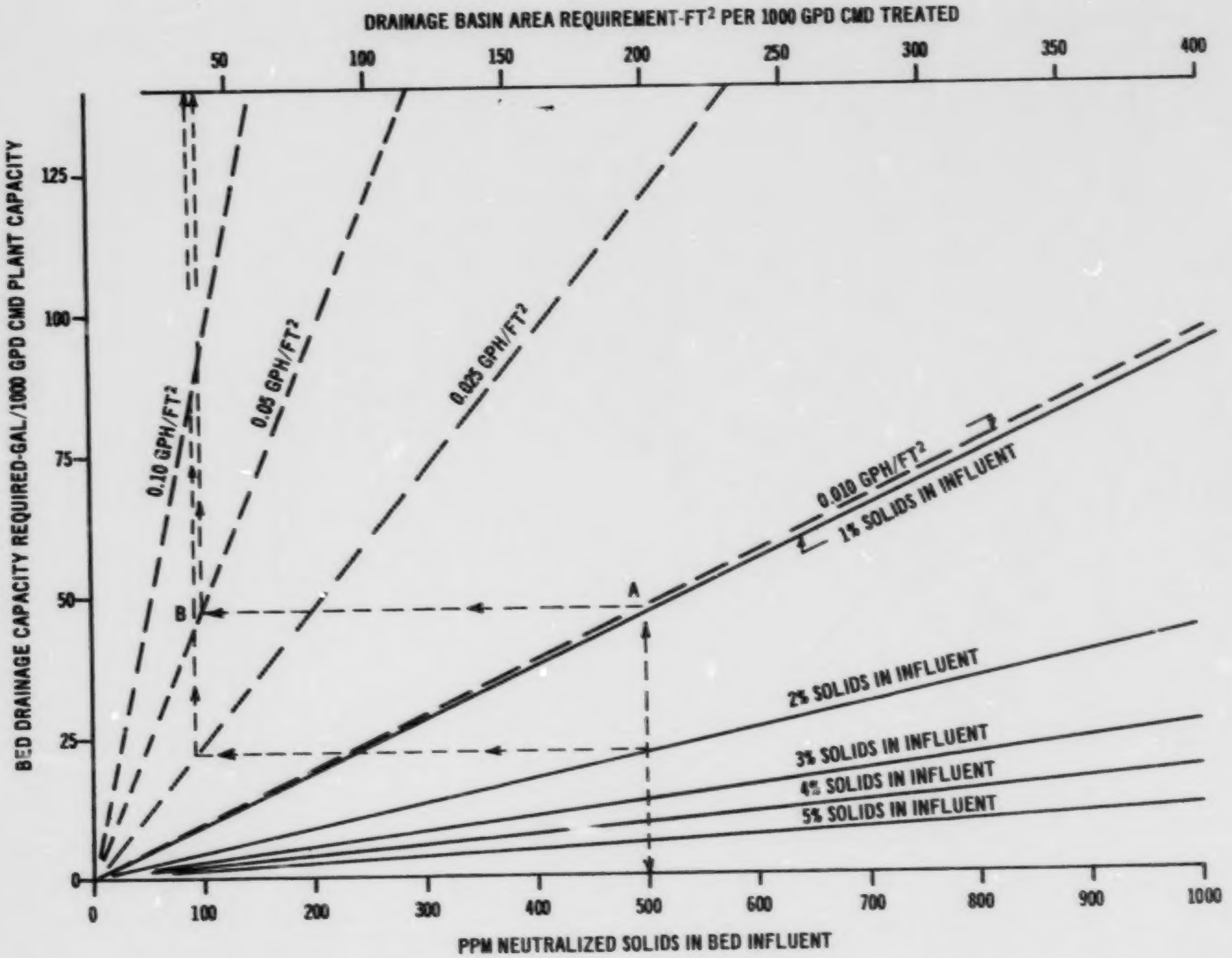


Figure 63. Sludge drying basin area calculator

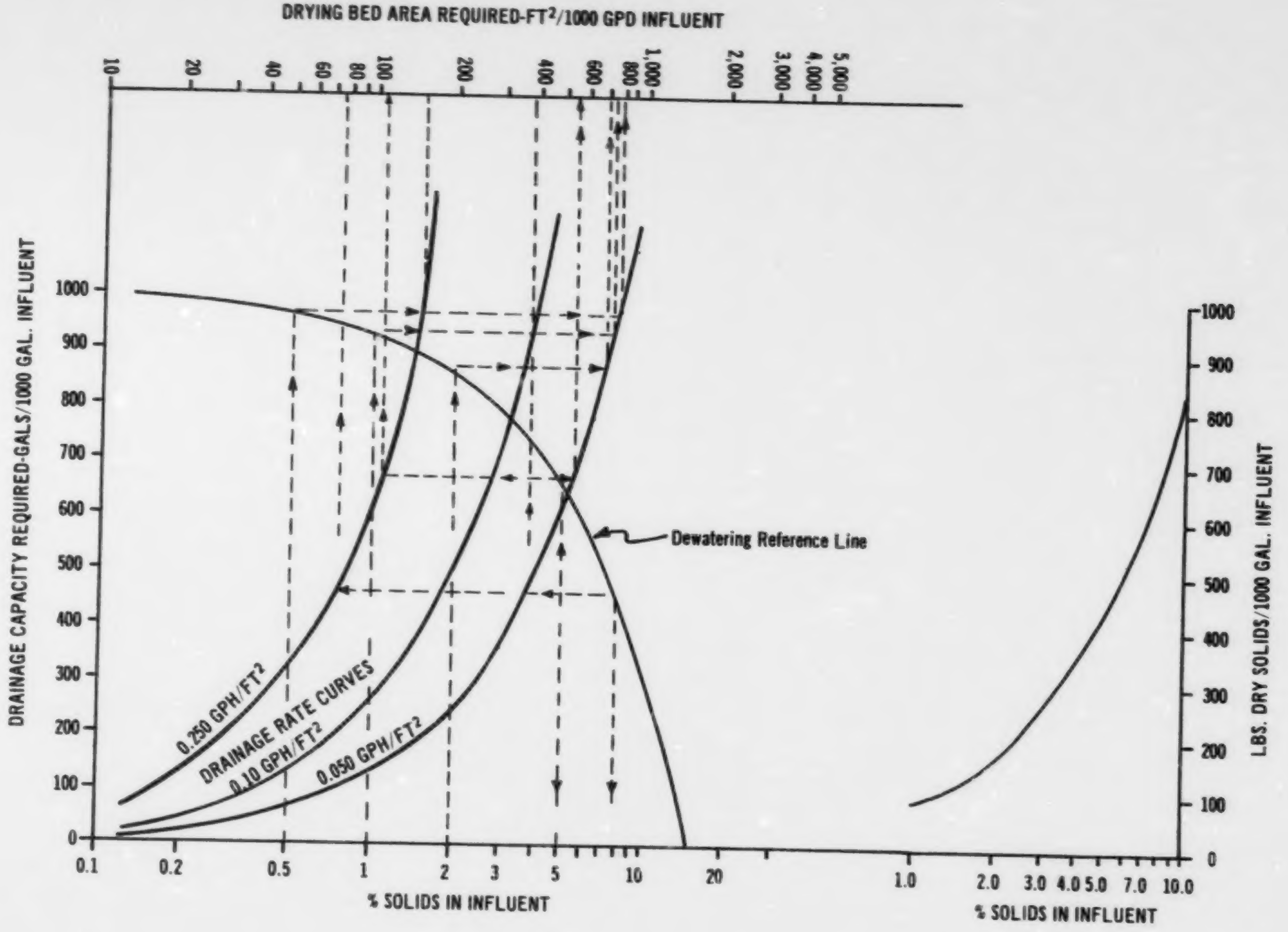


Table 66. TYPICAL SLUDGE DRYING BED OPERATIONAL PARAMETERS

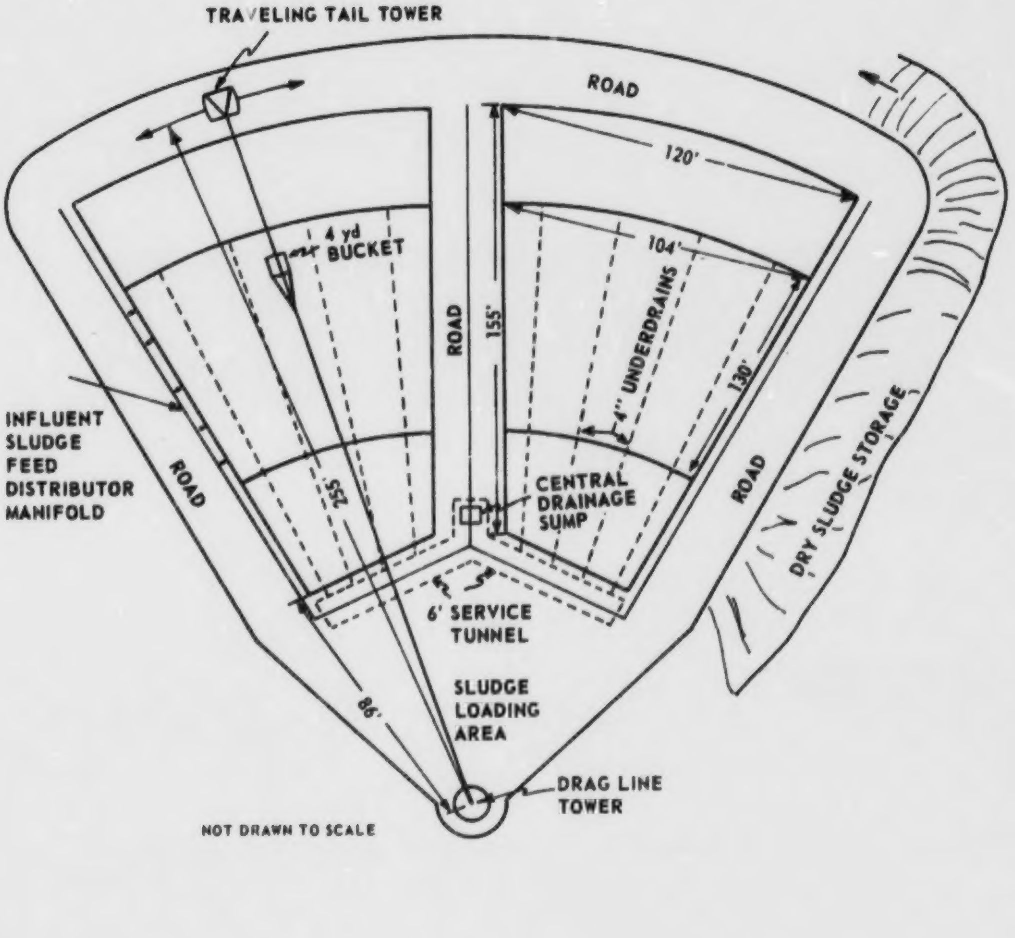
A	B	C	D	E	F	G	H	I	J	K	L
Basis: 2,000 ppm neutralized suspended solids settled to 2% solids slurry. 72 ft^2 drainage bed area required/1000 GPD raw AMD at sludge drainage rate of 0.05 GPH/ft^2.											
500.	36.	8.3	6.	808.	108.	4.	134	10	40	15	0.75
1,000.	72.	16.6	12.	1,616.	216.	8.	134	10	80	20	2.0
2,000.	144.	33.2	24.	3,232.	432.	16.	134	10	160 (80)	20	4.0
5,000.	360.	83.0	60.	8,080.	1,080.	40.	134	10	400 (200)	35	17.5
Basis: 1,000 ppm neutralized solids settled to 2% solids. 42 ft^2 drainage bed area required/1000 GPD raw CMD - 0.05 GPH/ft^2 drainage rate.											
500.	21.	(1) 4.2	(2) 3.	472.	63.	2.3	158	10	24	13	0.65
1,000.	42.	8.3	6.	944.	126.	4.7	158	10	48	16	1.60
2,000.	84.	16.6	12.	1,880.	252.	9.4	158	10	96 (48)	16	3.20
5,000.	210.	41.5	30.	4,720.	630.	23.4	158	10	240 (120)	25	12.50

A = CMD Plant Capacity - GPD x 10^3.
B = Drainage Bed Area Required - $ft^2 \times 10^3$=(A/1000 x 72)
C = Dry Weight of Solid Precipitated - lbs/day x 10^3 = (A x 8.34 x 2000)/10^6.
D = Volume of 15% Sludge Generated - GPD x 10^3 = C/0.15 x 9.25.
E, F, G = Volume of Sludge in bed 3 ft deep.
  E - Gallons x 10^3 = 7.48F.
  F - $ft^3 \times 10^3$ = 3B
  G - $yd^3 \times 10^3$ = F/27

H, I, J, K = Operating Schedule
  H = Fill Time - Days = E/D
  I = Drain Time - Days = E/2 x 1.2B
  J = Clean-out Time - Hours = G/100 TPH
  K = Down Time - Days = I + J
L = Excess Settling Basin Capacity Required-gallons x 10^6 = A x K/10

1. A/1000 x 42
2. A x 8.34 x 1000/10^6
3. Weight of 1 gallon of 15% sludge = 8.34 x 1.11 (5.6) = 9.25 lbs.

Figure 64. Suggested layout for dual sludge drainage basin system


TRAVELING TAIL TOWER
ROAD
120'
104'
4 yd BUCKET
ROAD
155'
4" UNDERDRAINS
130'
INFLUENT SLUDGE FEED DISTRIBUTOR MANIFOLD
ROAD
255'
CENTRAL DRAINAGE SUMP
ROAD
DRY SLUDGE STORAGE
6' SERVICE TUNNEL
SLUDGE LOADING AREA
86'
DRAG LINE TOWER
NOT DRAWN TO SCALE

the truck had to be available continuously in handling sludge from the filter or centrifuge or a temporary storage developed which required multiple handling.

The dewatered sludge was trucked to an abandoned "borrow pit" (several acres) adjacent to the Treatment Facility which would have capacity for years of plant operation at design capacity. The drying and volume reduction of the dewatered sludge continued slowly in the disposal area by percolation and evaporation whether in warm or sub-freezing weather. The volume reduction as a result of freezing was previously detailed (about 75 percent). The volume reduction from drying is illustrated in Figure 65 for a dewatered limestone sludge. The small cylinder ($8.95\ in^3$) was utilized to separate a dewatered specimen (left) which was dried in the laboratory. The final cylinder volume was $1.76\ in^3$ (right) for a reduction of about 80 percent.

After the dumped sludge reached a moisture content near that of the surrounding soil, it was graded, compacted, and seeded with common grasses, which germinated and grew. The seeding prevents possible windage of the dried sludge.

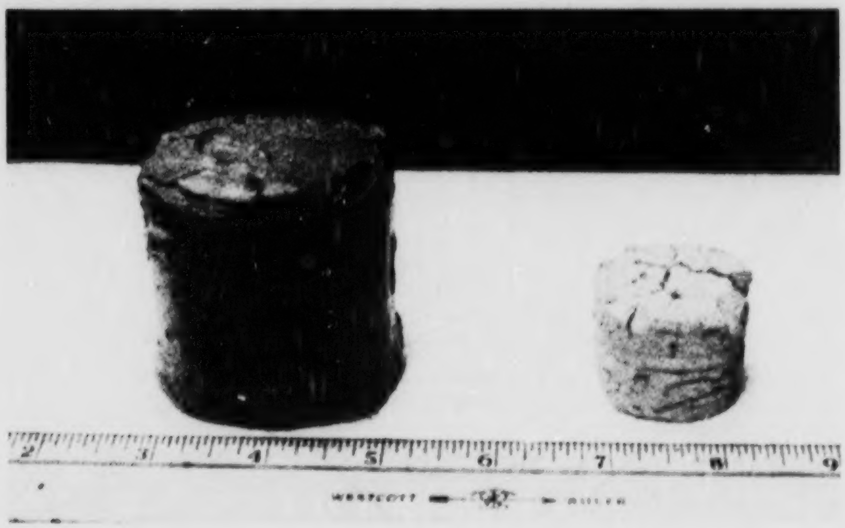


Figure 65. Volume Reduction in Dewatered Sludge When Dried

SECTION XV

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SECTION XVI

GLOSSARY

Air Oxidation - A process in which a substance, as ferrous sulfate, is oxidized to a higher oxidation number by reaction with oxygen contained in air.

Attrition - An autogenous process which maintains surfaces of solids clean and free of foreign substances, as reaction products. It results from frictional action upon contact of two particles.

Autotrophic - Needing only inorganic compounds for nutrition.

Biochemical Oxidation - A process in which a substance, as ferrous sulfate, is oxidized to a higher oxidation stage by some mechanism in which the metabolic life process predominates.

Calcined - A material which has been subjected to very high temperatures which changes its composition. As calcined lime, prepared by heating limestone with the subsequent removal of carbon dioxide.

Cation Exchange Reaction - The replacement of undesirable cation components in coal mine drainage by more acceptable soluble ions by solubility principles.

Clarifier - A mechanical device used in solid-fluid separation of suspensions designed to provide a clear, particle-free liquid product.

Coal Mine Drainage (CMD) - A water transferred from a coal mine environment. Such waters usually have enhanced dissolved solids and poor quality.

Compression - A condition which develops during quiescent particle sedimentation in which the particle-liquid structure becomes firm and develops compressive strength. The layer of solids supports layers above it, reducing subsidence by mechanical support propagated upward from the bottom of the containing vessel.

Densator - A proprietary type of thickener-reactor.

Dewaterable Sludge - An aqueous suspension of sparingly insoluble materials from which a major portion of the water may be removed by physical processes.

Dynamic System - One which is continuous by the addition of reactants and removal of products.

Effluent - A water product leaving a treatment process or system.

Feed or Raw Water - A water entering a treatment process of system. It is usually applied to those waters entering the first stage of a treatment system.

Gel - A process in which particulate matter and associated water molecules interact to form a physical state which resists flow.

Heterotrophic - Obtaining nourishment primarily from organic matter.

Influent - A water entering a treatment process.

Mineralized Waters - Waters which contain soluble mineral substances acquired by the water as it passed through strata of the earth.

Neutralization - A chemical reaction or processes which decreases the hydrogen ion content of coal mine drainage.

Neutralization Equivalent - The quantity of a chemical needed to react completely and without excess in a neutralization reaction. The quantity is consistent with that defined stoichiometrically by a balanced chemical equation.

Polluted Water - Waters containing substances inimical to normal utilization and which are not indigenous to most waters. The term as used is not restricted to biological contamination.

Rat Holing - A phenomena observed in settled sludges during removal of sludge by a vacuum line. The sludge tends to move more slowly than the clarified water and does not readily disperse. An interface envelope forms such that water, not sludge, is removed by the pump.

Reactivity - The extent of response, usually chemical, between two or more materials.

Red Dog - A non-volatile combustion product of the oxidation of coal or coal refuse. Most commonly applied to material resulting from in situ, uncontrolled burning of coal or coal refuse piles. It is similar to coal ash.

Secondary Sludge Dewatering - A subsequent process to solid-fluid separation in which more water is removed from the solids.

Settleability - An arbitrary concept for the behavior of solids sedimenting in a fluid which seeks to describe a variety of settling characteristics.

Slaking - A hydration process which converts calcined lime to hydrated lime.

Sludge - A thick, aqueous suspension of sparingly insoluble materials. Usually, but not necessarily, composed of valueless, waste substances.

Sludge Blanket - An interface between clear supernatant and settling solids in a clarifier or thickener. It generally represents the transition zone.

Slurry - A dilute aqueous suspension of a sparingly insoluble material.

Supernatant - The clear liquid which occurs above a sludge in a sedimentation operation.

Surface Biochemical Reactor - A device providing means by which biochemical reactions are carried out on the surface of an inert substrate.

Thickener - A mechanical device used in solid-fluid separations of suspensions designed to concentrate or thicken the solids.

Thixotropic - A rheological property of gelled sludges describing an increased tendency to flow after stirring or mixing.

Transition Zone - A region of sedimentation process in which free settling particles approach and contact each other to exist in a constrained structure so that all settle at the same rate.

Unwanted Reactions - A secondary or side reaction which consumes reagents or results in products beyond those desired, or planned, by an arbitary definition.

Vat Reactor System - A means by which chemical or biochemical reactions are carried out submerged in a liquid held in a large volume container.

Yellowboy Process - A conventional method of treating coal mine drainage by addition of hydrated lime, followed by ferrous iron oxidation by oxygen in air, and the separation of insoluble impurities by sedimentation.

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SECTION XVII

APPENDIX A

DESCRIPTION OF WATER CHARACTERISTICS AT PLANT SITE

The site of the Hollywood Facility was selected through the former Department of Mines and Mineral Industries and other cooperating Pennsylvania State agencies. After a review of several potential locations, the New Shawmut Mining Company (Mr. A. Palumbo, President), St. Marys, Pennsylvania, made the 40 acre tract available. It is in Huston Township, Clearfield County, near the Village of Hollywood.

This site, adjacent to Pennsylvania Route 255, includes coal mine drainage flows from two non-working deep mines, Proctor No. 1 and Proctor No. 2, from a small stream (Tyler Run) related to these mines and associated abandoned surface mines and from Bennett's Branch. The latter is a stream normally 60 feet wide, which receives drainage from several outfalls upstream. Bennett's Branch would insure a reliable water source for the Facility through all seasons. These streams pass through the site and flow via Bennett's Branch and the Sinnemahoning to the West Branch of the Susquehanna River.

The water quality of these streams available at the Facility have been monitored monthly between the springs of 1966 and 1971. The analyses are summarized in Table 1. A more detailed analysis of the waters in January 1968 is given in Table 2. The monthly samples were taken at stream locations where the waters are collected for transfer to the Treatment Facility. In addition, during the active portion of the study, two further points down stream from the Facility, referred to as Stream Combinations No. 1 and No. 2, were sampled. The former combination would include the flow from Proctor No. 2, Proctor No. 1, Tyler Run and also the treated effluent from the Facility. The second combination was further downstream in Bennett's Branch, and would represent the combined water of all four sources, including any plant effluent. There was a variable overflow through a borehole, about one mile northwest of the Facility originating in the Proctor No. 2 mine. This source was also monitored.

The seasonal variation in the water quality of these streams is shown in Figures 1 through 5. All the waters reflect change from precipitation and subsequent run-off. The seasonal changes are most pronounced in the stream samples.

Table 1. SUMMARY STREAM ANALYSIS AT HOLLYWOOD SITE
June, 1966 - June, 1971

	Proctor No. 2	Proctor No. 2 Bore Hole	Proctor No. 1	Tyler Run	Bennetts Branch	Stream Combinations No. 1	Stream Combinations No. 2
				pH			
Average	2.98	3.20	3.34	3.43	3.88	3.26	3.36
High	3.81	3.41	4.40	4.30	5.10	3.71	4.10
Low	2.69	2.85	2.60	2.91	2.80	2.79	2.88
			ACIDITY (mg/L $CaCO_3$)				
Average	3302	2485	489	160	164	479	288
High	8112	4121	1271	420	523	1138	584
Low	478	1718	48	54	15	123	58
			IRON II (mg/L)				
Average	548	804	83	4	29	84	38
High	2150	1579	199	90	105	290	116
Low	23	509	11	<1	2	1	5
			IRON (TOTAL) (mg/L)				
Average	780	809	161	40	52	86	39
High	5728	1579	2538	1269	1340	295	117
Low	23	543	10	1	5	8	10

Table 1 (continued). SUMMARY STREAM ANALYSIS AT HOLLYWOOD SITE
June, 1966 - June, 1971

	Proctor No. 2	Proctor No. 2 Bore Hole	Proctor No. 1	Tyler Run	Bennetts Branch	Stream Combinations	
						No. 1	No. 2
			ALUMINUM (mg/L)				
Average	245	232	87	32	25	100	47
High	861	393	229	131	87	528	115
Low	78	184	13	1	0	21	10
			CALCIUM (mg/L)				
Average	201		96	59	36		
High	368		144	108	56		
Low	60		40	20	20		
			MAGNESIUM (mg/L)				
Average	97		76	64	40		
High	145		132	114	46		
Low	3		10	28	33		
			MANGANESE (mg/L)				
Average	8.2	9.5	2.6	3.0	3.5	4.0	2.9
High	9.0	12.0	3.5	6.9	9.4	7.3	5.7
Low	6.8	6.7	1.3	1.2	1.2	1.8	1.3
			SULFATE (SO_4^{-2}) (mg/L)				
Average	2587	3295	809	301	260	1156	424
High	7491	5276	1188	712	747	2018	1009
Low	815	2650	107	118	48	229	101

Table 1 (continued). SUMMARY STREAM ANALYSIS AT HOLLYWOOD SITE
June, 1966 - June, 1971

	Proctor No. 2	Proctor No. 2 Bore Hole	Proctor No. 1	Tyler Run	Bennetts Branch	Stream Combinations	
						No. 1	No. 2
			CONDUCTIVITY (MICROMHOS)				
Average	2666	3311	1223	602	519	1150	704
High	3090	5420	1620	1070	1335	2320	1550
Low	2348	2580	194	270	132	5	5
			DENSITY (20°C)				
Average	1.00505	1.0058	1.00305	1.0021	1.0027	1.0031	1.0032
High	1.0060	1.0080	1.0040	1.0031	1.0041	1.0050	1.0050
Low	1.0041	1.0045	1.0020	1.0000	1.0010	1.0010	1.0012

Table 2. DETAILED STREAM ANALYSES AT HOLLYWOOD SITE
COLLECTION DATE JANUARY 11, 1968

Constituent	Proctor No. 2	Proctor No. 2 Bore Hole	Proctor No. 1	Tyler Run	Bennetts Branch
pH	3.06	3.40	3.50	3.58	3.80
Acidity, mg/l CaCO^{c}	478	2810	469	124	210
Dissolved solids, mg/l	1324	5892	1342	429	545
Fe (total), mg/l	23	911	84	<1	49
Fe (II), mg/l	23	911	83	<1	49
Al, mg/l	78	229	91	33	36
Ca, mg/l^{b}	-	185	-	-	-
Mg, mg/l	-	122	-	-	-
SO_4, mg/l	815	3457	827	266	322
Na, mg/l^{a}	49	34	19	4	7
K, mg/l^{a}	0.8	7	2	0.7	1
Mn, mg/l	10.3	10.6	2.1	1.5	3.3
Ni, mg/l^{b}	0.7	2.9	0.6	0.1	0.2
V, mg/l^{b}	0.009	0.2	0.005	0.002	0.02
Cu, mg/l^{b}	0.1	0.3	0.04	0.03	0.2
Mo, mg/l^{b}	ND	ND	ND	ND	ND
Zn, mg/l^{b}		5.9			
Si, mg/l^{b}		55			
Sr, mg/l^{b}		3.4			
Ba, mg/l^{b}		0.6			
Ti, mg/l^{b}		2.0			
Hg, ppb^{c}		<50			
Total dissolved solids defined, mg/l	978	5032	1026	308	419

Table 2 (continued). DETAILED STREAM ANALYSES AT HOLLYWOOD SITE COLLECTION DATE JANUARY 11, 1968

Constituent	Proctor No. 2	Proctor No. 2 Bore Hole	Proctor No. 1	Tyler Run	Bennetts Branch

[a] Flame photometer, on raw water.

[b] Emission spectroscopy from evaporated residue (105°C).

[c] Atomic absorption after concentration on gold, N.H. Suhr, Pennsylvania State University, University Park, Pa.

Figure 1. Seasonal variation in water quality at Hollywood Facility - Proctor No. 2

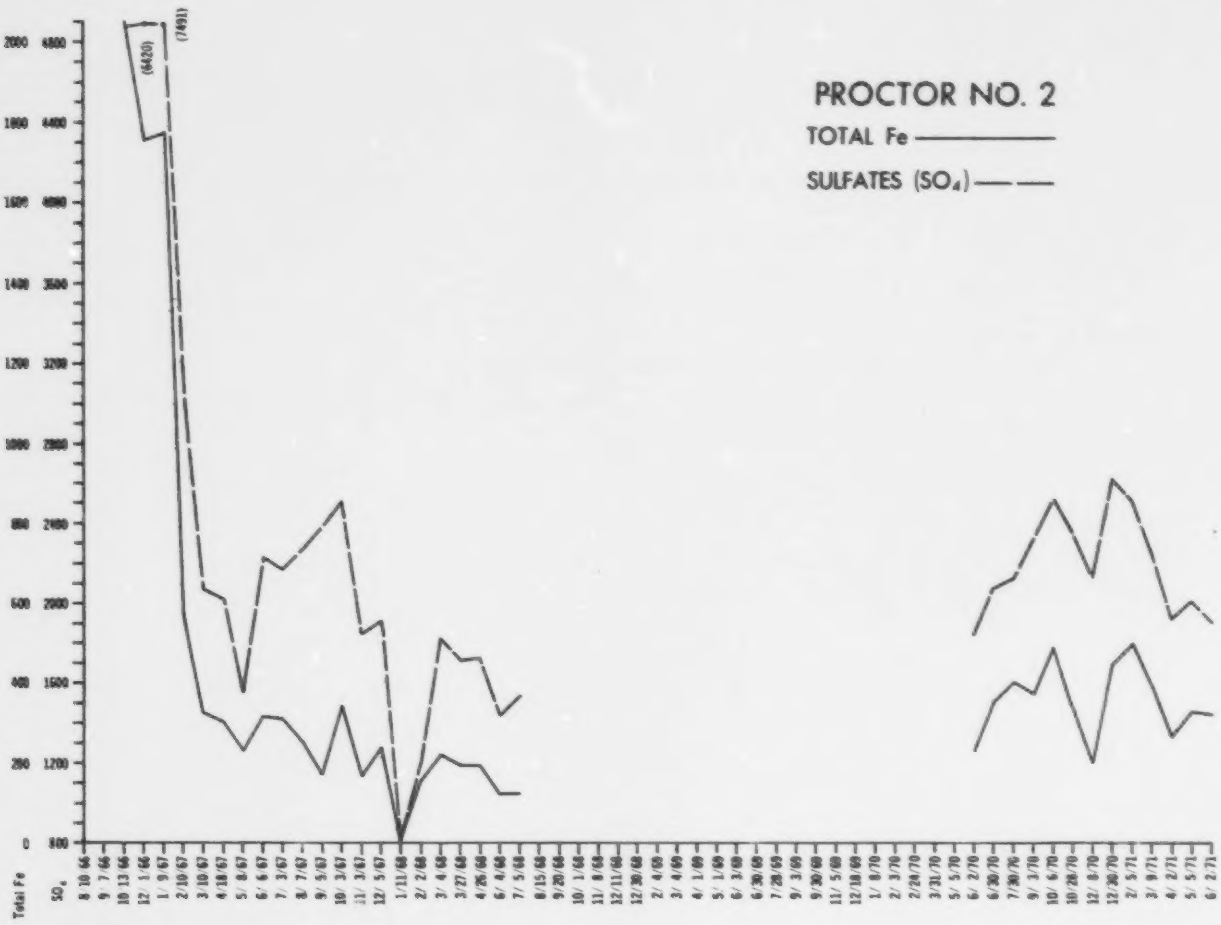


Figure 1 (continued). Seasonal variation in water quality at Hollywood Facility - Proctor No. 2

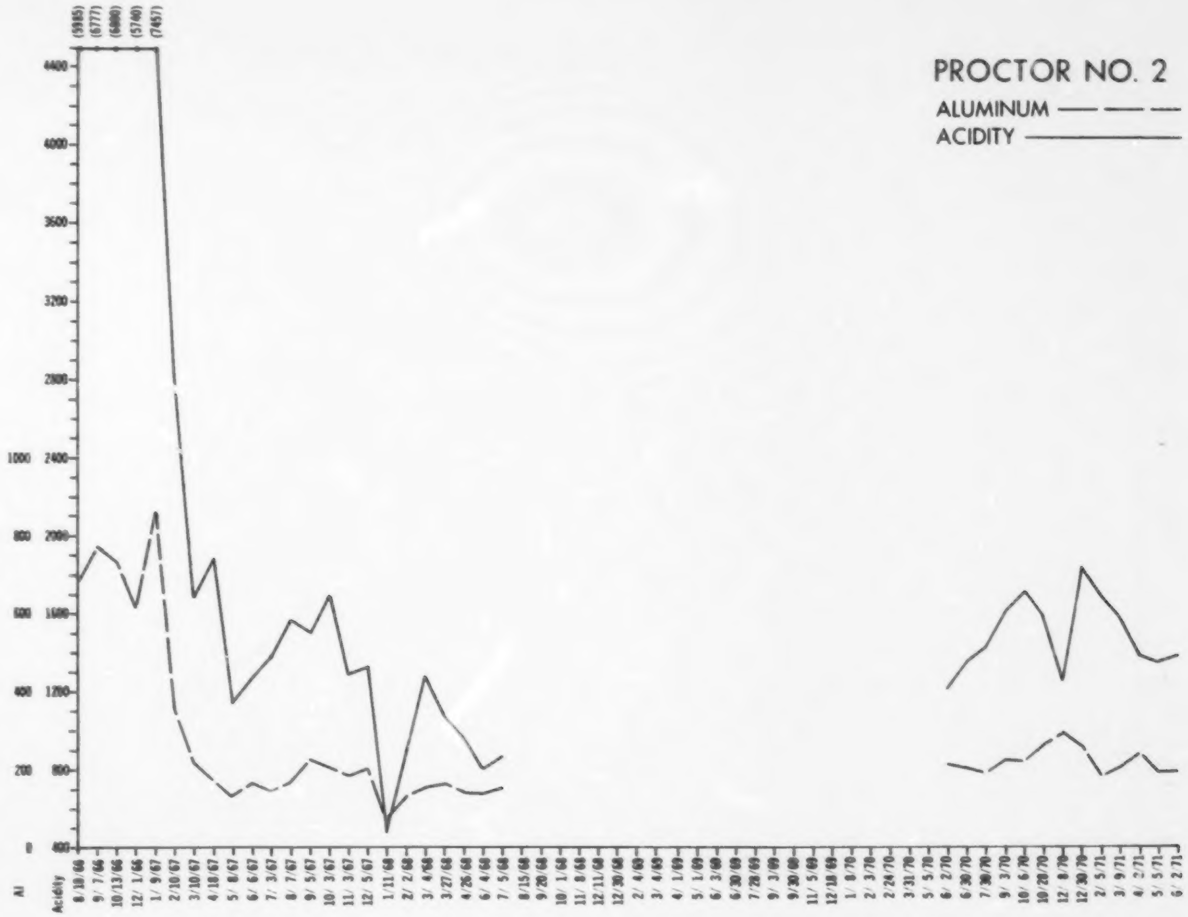


Figure 2. Seasonal variation in water quality at Hollywood Facility - Proctor No. 2 bore hole

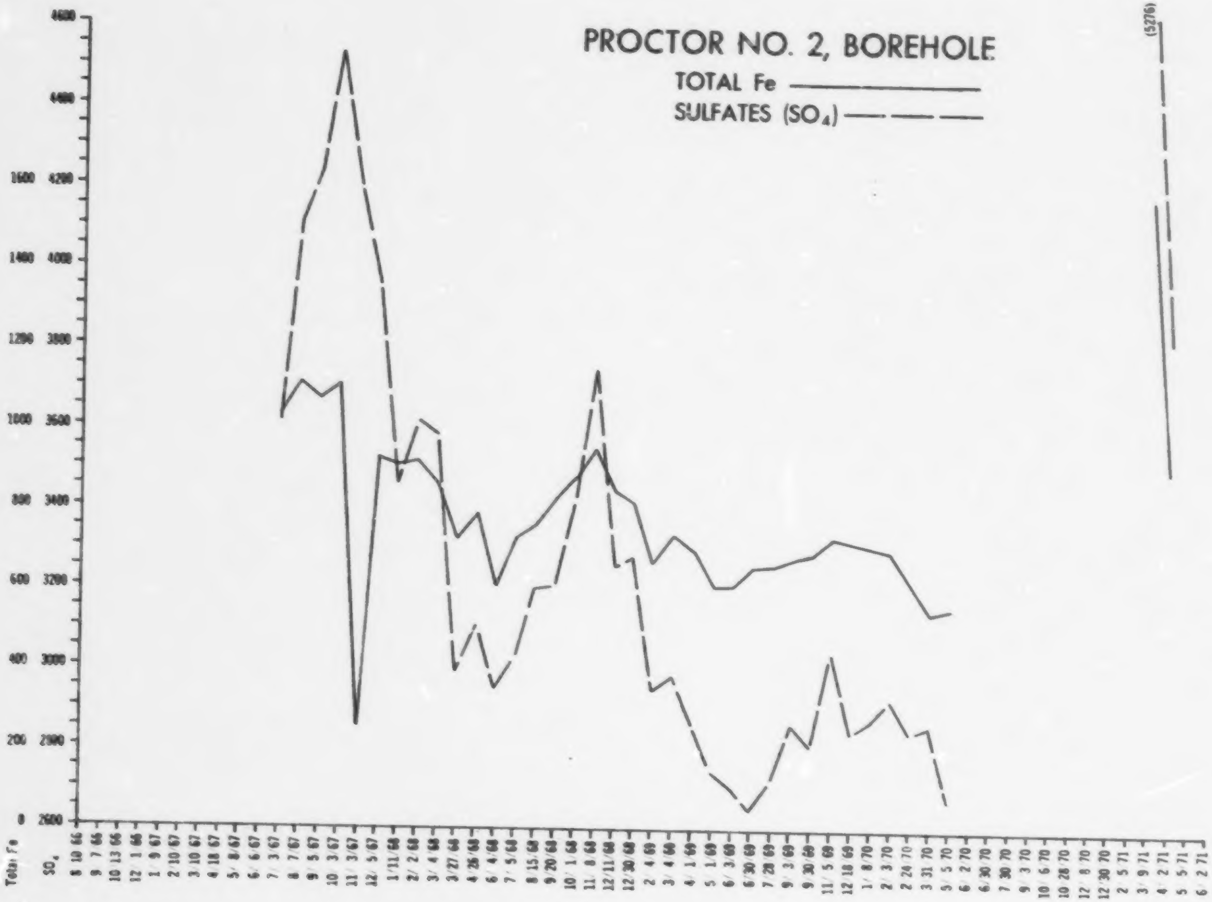


Figure 2 (continued). Seasonal variation in water quality at Hollywood Facility - Proctor No. 2 bore hole

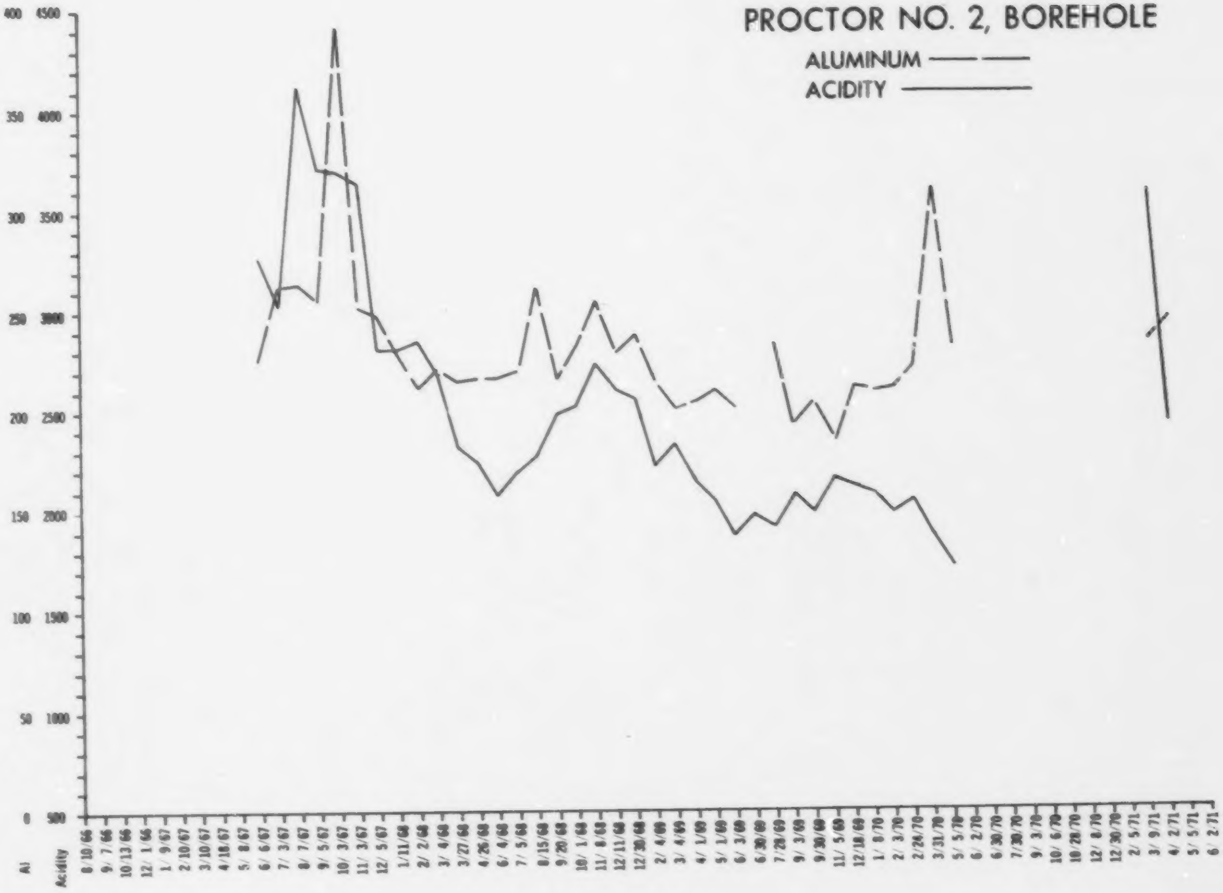


Figure 3. Seasonal variation in water quality at Hollywood Facility - Proctor No. 1

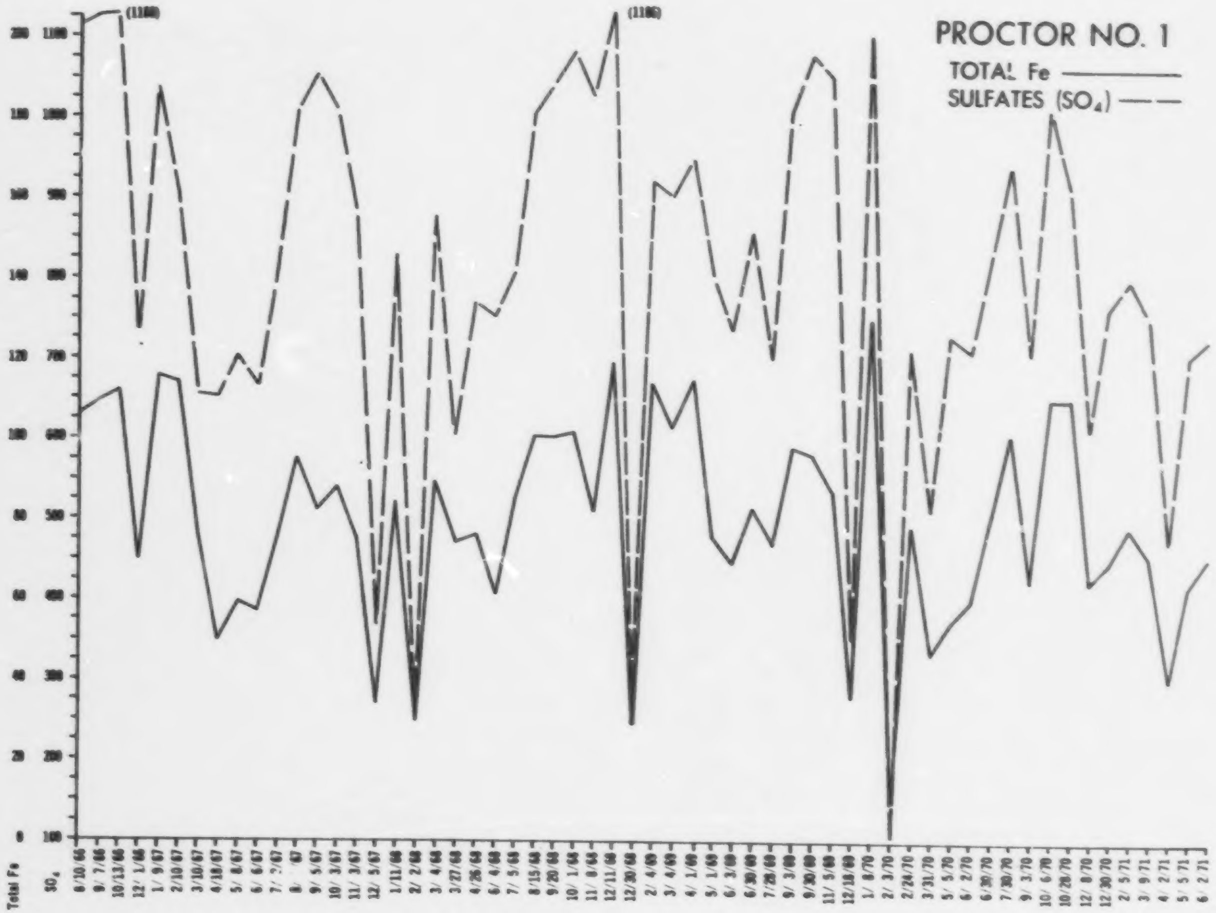


Figure 3 (continued). Seasonal variation in water quality at Hollywood Facility - Proctor No. 1

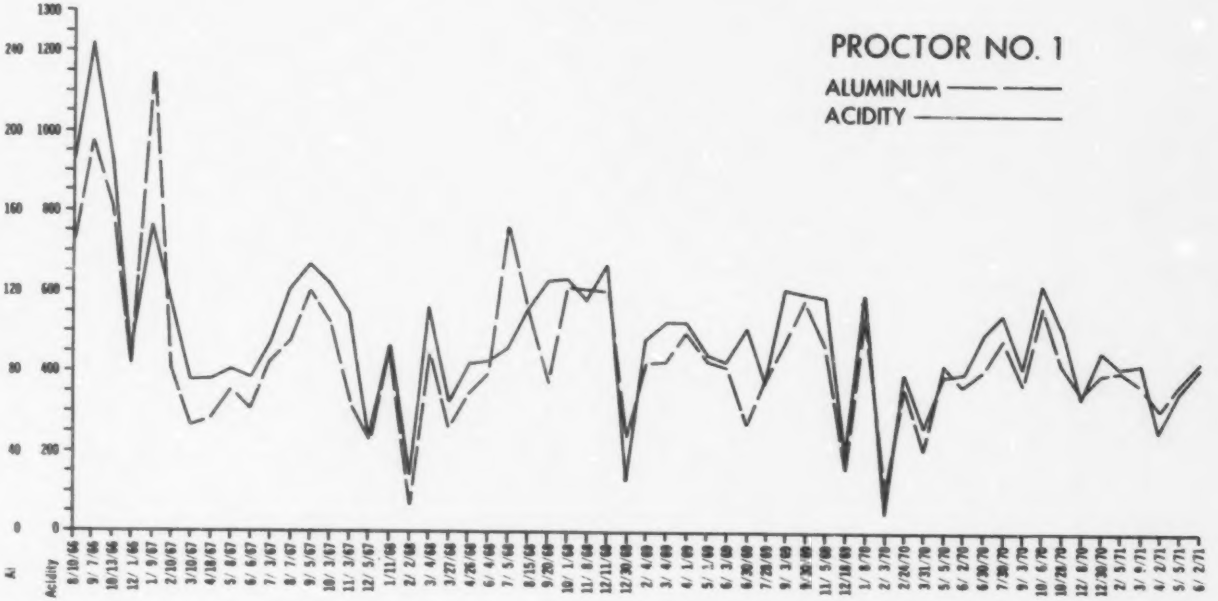


Figure 4. Seasonal variation in water quality at Hollywood Facility - Bennett's Branch

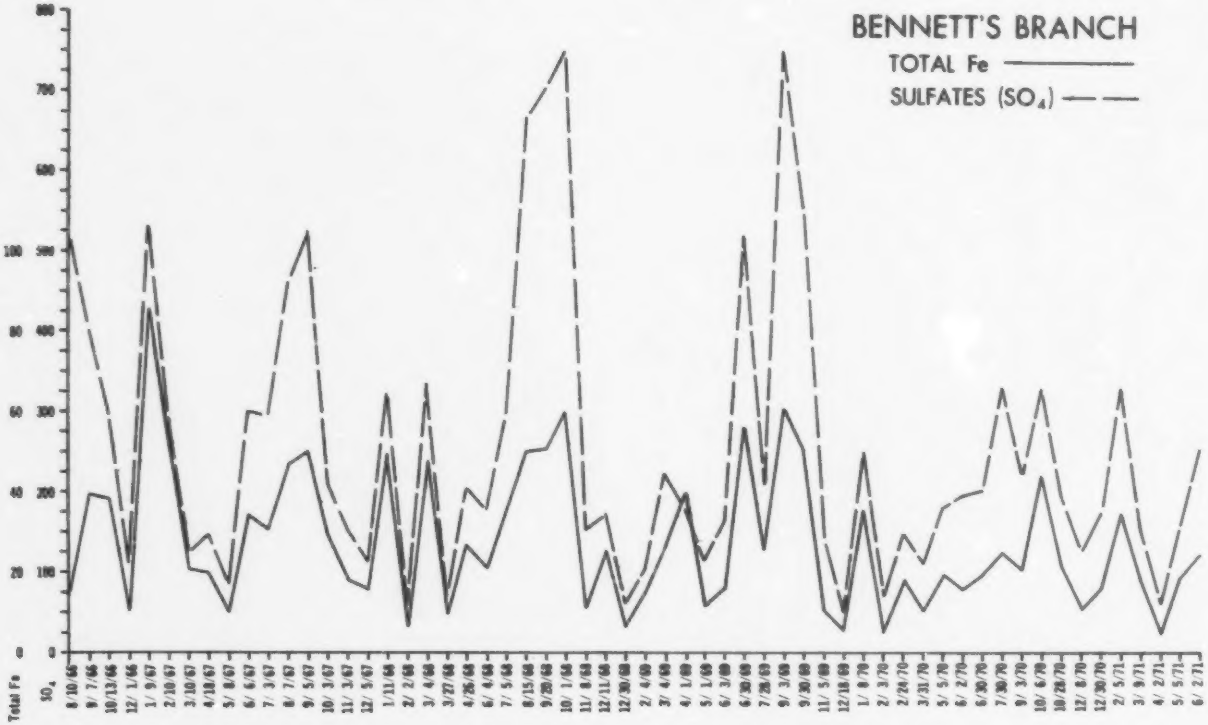


Figure 4 (continued). Seasonal variation in water quality at Hollywood Facility - Bennett's Branch

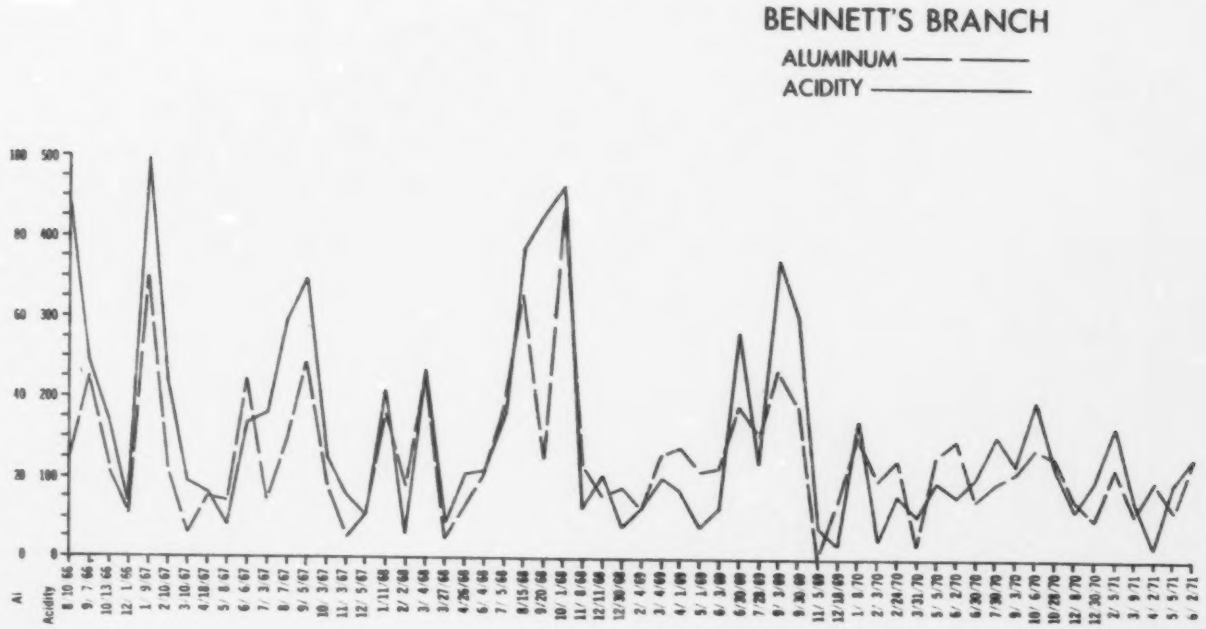


Figure 5. Seasonal variation in water quality at Hollywood Facility - Tyler Run

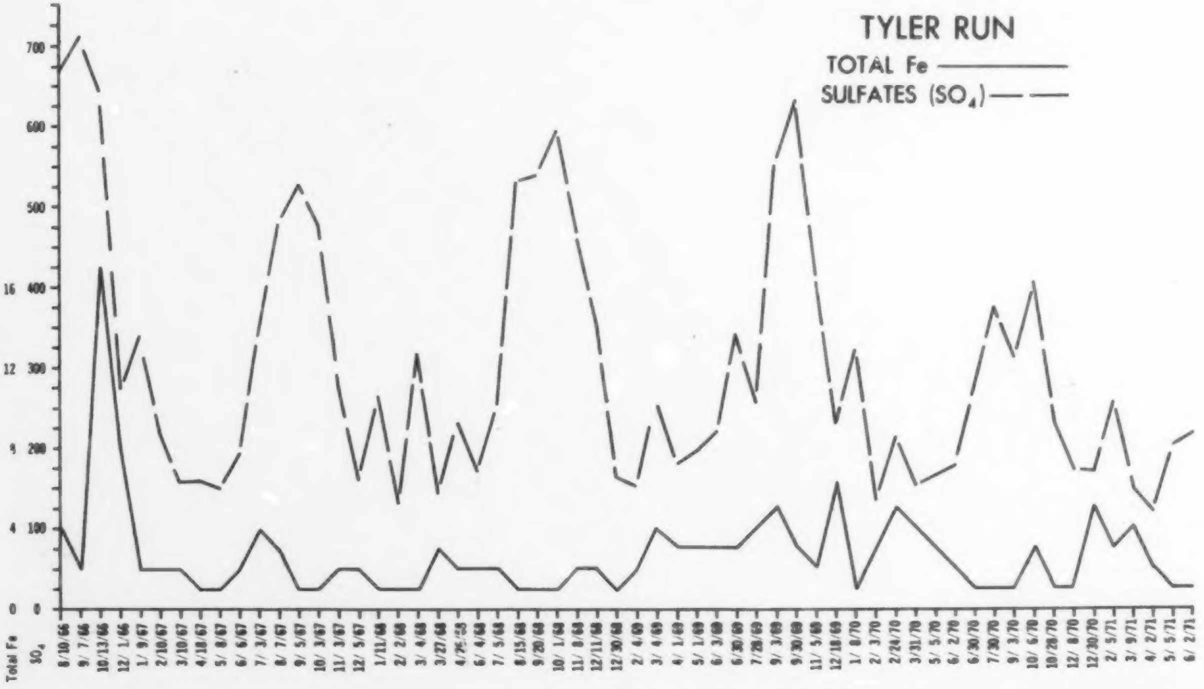
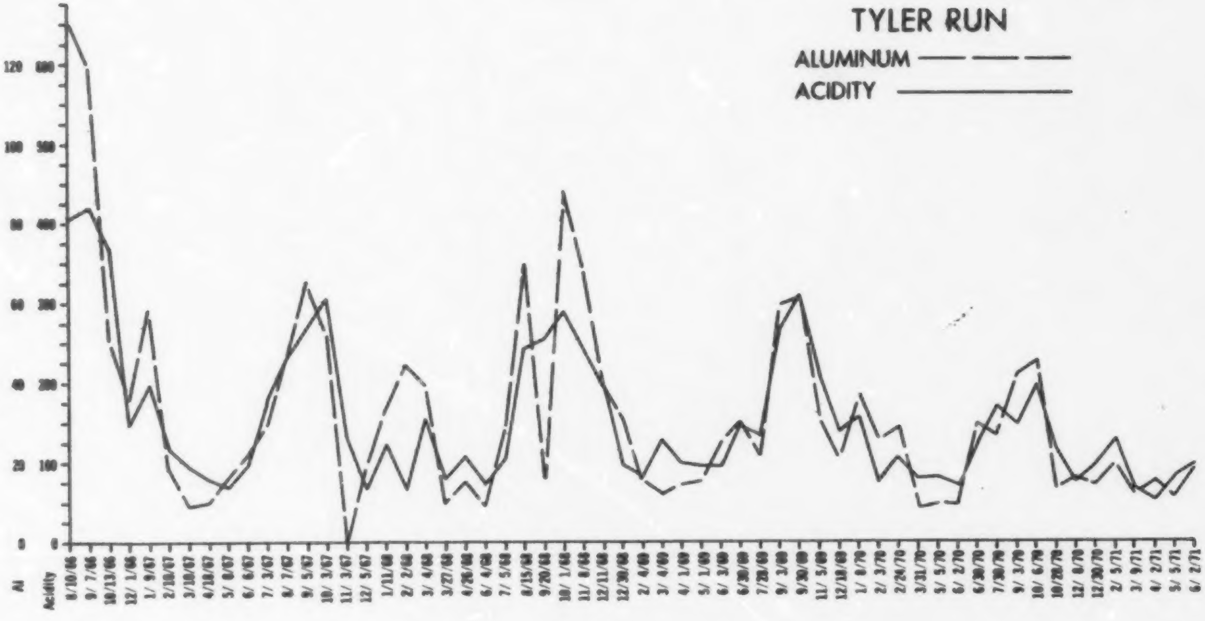


Figure 5 (continued). Seasonal variation in water quality of Hollywood Facility - Tyler Run



APPENDIX B

PLANT DESIGN

Table 1. UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Major Collection and Treatment Units

Three (3) pumping stations
Holding lagoon
Flash mixer
Oxidation unit
Rock filter
Densator unit
Settling lagoon
Drum sludge vacuum filter
Limestone neutralization unit
Lime, sodium hydroxide and sodium carbonate feed units
Polyelectrolyte feed unit
Sludge drying lagoon
Control building
Garage and storage unit

Control Building

Construction
  Reinforced concrete sub-structure, concrete block walls, steel frame, concrete plank roof, and rear loading platform

Units and facilities
  Electrical controls
  Pipe gallery
  Rate-of-flow control
  Flow, pH, temperature, iron, turbidity and DO indicators and recorders
  Flash mixer
  Two (2) blowers and appurtenant equipment

(1) Based on information furnished by the University; subject to change as additional information is made available
  Densator unit influent sump and pumps, sludge pumps, and appurtenant equipment
  Drum sludge vacuum filter, cake conveyor, coal storage and feed facilities, conditioning tank, and appurtenant equipment

Table 1 (continued) UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Hydrated lime storage bin, hydrated lime feed facilities, and apurtenant equipment
Quicklime storage bin, slaker and other feed facilities, and appurtenant equipment
Sodium hydroxide storage and feed facilities, and appurtenant equipment
Sodium carbonate storage and feed facilities, and appurtenant equipment
Polyelectrolyte storage and feed facilities, and appurtenant equipment
Plant water pumps and piping
Office
Laboratory
Workshop
Conference room
Kitchen
Sanitary facilities, and septic tank and tile field
Sleeping quarters

Possible Future Construction

Pre-coat vacuum filter
Recirculate rock filter and/or Densator unit effluent to rock filter influent line
Rock filter cover
Water storage tank

Proctor No. 1 Pumping Station

Pump suction downstream from confluence of 2 mine drainage discharges originating on north side of U.S. Route 255

Point of withdrawal from stream	Concrete sump set in bed of stream, shape stream channel as needed
Sump capacity	1,000 gallons
Type of pump housing structure	Below grade, concrete, acid resistant, locate top above 1200 elevation

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Pump	
Type	Centrifugal, constant speed with by-pass for varying delivery
Capacity	350 gpm
Construction	Acid resistant
Suction	Flooded
Discharge	Control building pipe gallery and holding lagoon
Piping	PVC
Miscellaneous	
Flow indicator; automatic pump stop; dry well sump and pump; dry well electric unit heater and ventilation; electrical controls (duplicate on-off control in control building)	
Proctor No. 2 Pumping Station	
Pump suction from Proctor No. 2 and Tyler Run	
Point of withdrawal from streams	Concrete sumps set in bed of both streams, shape stream channels as needed
Sump capacity - each	1,000 gallons
Type of pump housing structure	Below grade, concrete, acid resistant, locate top above 1205 elevation
Pumps	
Number	2 (1 for Proctor No. 2 and 1 for Tyler Run)
Type	Centrifugal, constant speed with by-pass for varying delivery
Capacity - each	350 gpm
Construction	Acid resistant
Suction	Flooded
Discharge	Control building pipe gallery and holding lagoon

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Piping	PVC
Miscellaneous	
Flow indicators; automatic pump stops; dry well sump and pump; dry well electric unit heater and ventilation; electrical controls (duplicate on-off controls in control building)	
Bennett Branch Pumping Station	
Pump suction from Bennett Branch	
Point of withdrawal from stream	Portion of Bennett Branch diverted from main channel to concrete sump located by shoreline, shape stream channel as needed
Sump capacity	1,000 gallons
Type of structure	Below grade, concrete, acid resistant, locate top above Bennett Branch flood level
Pump	
Type	Centrifugal, constant speed with by-pass for varying delivery
Capacity	350 gpm
Construction	Acid resistant
Suction	Flooded
Discharge	Control building pipe gallery and holding lagoon
Piping	PVC
Miscellaneous	
Flow indicator; automatic pump stop; dry well sump and pump; dry well electric unit heater and ventilation; electrical controls (duplicate on-off controls in control building)	

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Holding Lagoon	
Capacity	258,000 gallons
Dimensions	
Diameter	112 feet
Center depth (liquid level to bottom)	5.6 feet
Freeboard	2 feet
Side slopes	4:1
Construction	Earth and bentonite mixed, and placed in compacted layers
Inlet	Header to receive mine drainage from piping gallery and discharge tangentially at periphery; mechanical liner at discharge point
Outlet	Located at bottom center of lagoon protective screen
Piping	PVC
Miscellaneous Lagoon by-pass and overflow; piping arranged to flush outlet protective screen; rate-of-flow control on discharge line; influent and discharge to be metered (indicator-recorder-totalizer); continuous pH, temperature, iron and turbidity recording; drain to Tyler Run; effluent piping to flash mixer, Densator unit influent sump, and biological treatment units	
Flash Mixer	
Dimensions (L x W x SWD)	12 x 3 x 6.5 feet
Volume	1,750 gallons
Location of inlets	Bottom of unit
Location of outlets	Top of unit
Construction	Steel

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Mixers	2; portable so that position within unit can be varied
Piping	PVC
Miscellaneous	
Unit to be constructed so that detention time can be varied by inserting or removing 2 baffles located 4 feet from each end; provision for overflow from each compartment; bottom drain; unit to receive mine drainage from holding lagoon or directly from pipe gallery; effluent piping to oxidation unit and Densator unit influent sump	
Oxidation Unit	
Dimensions (L x W x SWD)	15 x 10 x 10 feet
Volume	11,200 gallons
Location inlets	Bottom of unit
Location of outlets	Top of unit
Construction	Steel, acid resistant
Turbine and surface aerators	
Oxygen transfer capabilities	1 - 30 lbs. O_2/hr.
Blowers	
Number	2
1 at	250 cfm
1 at	500 cfm
Air headers	PVC
Piping	PVC
Miscellaneous	
Unit to be constructed so that detention time can be varied by inserting or removing baffle located at mid-point; influent and piping to allow for use of each compartment; air to be supplied by diffusors, turbine aerators, and surface aerators; unit to receive mine drainage from pipe gallery, holding lagoon and flash mixer; effluent piping to Densator unit influent sump and settling lagoon; flow indicator on air lines; bottom drain; continuous pH and DO recording in unit (location of probes to be varied)	

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Rock Filter	
Dimensions (Diam. x SRD)	35 x 5 feet
Volume	4,800 $ft.^3$
Media	
Type	Silica rock
Size	1.5-2 inches
Piping	PVC
Rotary distributor	Acid resistant
Construction	Concrete, acid resistant
Miscellaneous Unit to receive mine drainage from pipe gallery and holding lagoon; effluent piping to Densator unit influent sump; make provision for future recirculation of rock filter effluent to influent line	
Densator Unit	
Dimensions	
Overall (Diam x SWD)	25 x 14 feet
Primary reaction zone (Diam. x SWD)	12 x 2.25 feet
Secondary reaction zone (Diam. x SWD)	12 x 8.50 feet
Clarification zone	
Surface area	378 $ft.^2$
SWD	10.5 feet
Volumes	
Primary reaction zone	1,900 gallons
Secondary reaction zone	7,200 gallons
Clarification zone	29,400 gallons
Construction	Steel, acid resistant
Influent pumps and sump	
Sump	
Capacity	1,000 gallons
Construction	Concrete, acid resistant
Overflow	Tyler Run

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Rock Filter	
Dimensions (Diam. x SRD)	35 x 5 feet
Volume	4,800 ft.^{3}
Media	
Type	Silica rock
Size	1.5-2 inches
Piping	PVC
Rotary distributor	Acid resistant
Construction	Concrete, acid resistant
Miscellaneous	
Unit to receive mine drainage from pipe gallery and holding lagoon; effluent piping to Densator unit influent sump; make provision for future recirculation of rock filter effluent to influent line	
Densator Unit	
Dimensions	
Overall (Diam x SWD)	25 x 14 feet
Primary reaction zone (Diam. x SWD)	12 x 2.25 feet
Secondary reaction zone (Diam. x SWD)	12 x 8.50 feet
Clarification zone	
Surface area	378 ft.^{2}
SWD	10.5 feet
Volumes	
Primary reaction zone	1,900 gallons
Secondary reaction zone	7,200 gallons
Clarification zone	29,400 gallons
Construction	Steel, acid resistant
Influent pumps and sump	
Sump	
Capacity	1,000 gallons
Construction	Concrete, acid resistant
Overflow	Tyler Run

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Pumps	
Number	2 (1 on stand-by)
Type	Centrifugal, manual variable delivery
Capacity (each)	35-450 gpm
Suction	Flooded
Construction	Acid resistant
Sludge pumps	
Number	2 (1 on stand-by)
Type	Positive displacement, manual variable delivery
Capacity (each)	10-125 gpm
Suction	Directly from bottom of unit; pumps located at outside bottom of unit
Construction	Acid resistant
Discharge	Vacuum filter, tank truck, sludge drying lagoon, oxidation unit influent, recycle into Densator unit or influent sump, settling lagoon
Piping	PVC

Miscellaneous

Unit to receive mine drainage from pipe gallery, holding laggon, rock filter and oxidation unit; effluent piping to plant water sump and Tyler Run, limestone neutralization unit, and flash mixer; flow indicators on influent and sludge lines; sample lines into Densator unit; bottom drain; scraping mechanism modified for thickening; continuous pH recording in Densator unit; make provision for future recirculation of overflow to rock filter influent line

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Settling Lagoon	
Construction	Earth placed in compacted layers, mechanical liner on bottom of flocculation and settling compartment
Freeboard	2 feet
Side slopes	2:1
Flocculation and settling compartment	
First section	
Volume	156,000 gallons
Dimensions (L x W x SWD)	86 x 52 x 8 feet
Bottom slope	3% toward influent end, 6% toward sludge withdrawal line
Inlet	Mechanical liner at discharge point
Second section	
Volume	63,000 gallons
Dimensions (L x W x SWD)	30 x 52 x 8 feet
Bottom slope	6% toward influent end, 6% toward sludge withdrawal line
Polishing compartment	
Volume	204,000 gallons
Dimensions (L x W x SWD)	70 x 70 x 8 feet
Sludge pump	
Type	Positive displacement, manual variable delivery
Capacity (each)	10-150 gpm
Construction	Standard
Suction	Flooded
Discharge	Vacuum filter, tank truck, sludge drying lagoon, oxidation unit influent
Type of structure	Concrete, acid resistant on outside

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Piping	PVC
Miscellaneous	
Sump and pump, electric unit heater, ventilation and electrical controls (duplicate on-off controls in control building) in pump structure; bottom drain to Tyler Run; discharge to be metered (indicator-recorder-totalizer); locate top above 1205 elevation; unit to receive mine drainage from oxidation unit, limestone neutralizer, and flash mixer; effluent piping to plant water sump and Tyler Run; flow indicators on sludge discharge lines	
Drum Sludge Vacuum Filter	
Filter	
Area	125 ft.^{2}
Dimensions (Diam. x L)	6 x 7 feet
Construction	Acid resistant
Type of drum discharge	Roll
Coal storage and feed	
Mesh	-60 to +80
Storage hopper	
Capacity	1,500 lbs.
Miscellaneous	
Hand loading from ground floor with 100 pound bags of pre-ground coal fines; level indicators; dust collector, vibrators; storage for 50-100 pound bags in control building, additional storage in garage	
Feeder	
Feed rate	
Maximum	500 lbs/hr
Minimum	20 lbs/hr
Type	Dry volumetric feed
Control	Manual variable delivery
Slurry mix tank and pumps	
Tank	
Capacity	100 gallons
Mixer	rapid
Pumps	
Number	2 (1 on standyby)

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Type	Positive displacement, manual variable delivery
Capacity (each)	2-20 gpm
Suction	Flooded
Construction	Acid resistant
Discharge	Conditioning tank
Conditioning tank	
Volume	200 gallons
Construction	Steel, acid resistant
Mixer	Slow
Discharge	Filter vat
Miscellaneous	
Conditioning tank to receive underflow from Densator unit and settling lagoon, polyelectrolyte slurry pump, and coal slurry pump	
Piping	PVC
Miscellaneous	
Filter wash and slurry water from plant water sump, or well; cake conveyor to dump truck for hauling to local strip pits or refuse piles; filtrate discharge to Tyler Run; float in filter vat to control start-stop position for sludge pump feeding conditioning tank, and coal slurry pump; float in slurry mix tank to control start-stop position of feeder and slurry water feed; flow indicators on slurry water feed and pump discharge; make provision for future pre-coat operation of filter	
Limestone Neutralization Unit	
Reactor	
Dimensions (Diam. x L)	5 x 20 feet
Degrees from horizontal	8
Limestone	
Size	0.25 inch
Volume	100 ft.^{3}
Construction	Steel, acid resistant
Miscellaneous	
Limestone added to upper end; mechanical rotation at 10 rpm for 2 minutes every 1/4 hour; reactor to receive mine drainage from Densator unit; fixed overflow pipe at lower end; provision for removal of spent limestone	

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Reactor sump	
Dimensions (L x W x SWD)	30 x 10 x 15 feet
Volume	11,200 gallons
Construction	Steel
Miscellaneous	
Bottom drain; effluent piping to settling lagoon	
Storage hopper	
Capacity	1,000 lbs.
Miscellaneous	
Loaded by bucket on hi-lift vehicle; level indicators	
Feeder	
Feed rate	
Maximum	80 lbs/hr
Minimum	2 lbs/hr
Type	Dry volumetric feed
Control	Manual variable delivery
Piping	PVC
Miscellaneous	
Storage hopper, feeder and upper end of reactor to be enclosed	

Hydrated Lime Storage and Feed

Type of hydrated lime	High calcium and dolomitic to be fed separately
Storage bin	
Capacity	25 tons
Miscellaneous	
Tank truck delivery (17-20 tons); blower hose; screw discharge to lime hopper; level indicators, dust collector; circulating fan; vibrators	
Lime Feeder	
Feed rate	
Maximum	500 lbs/hr
Minimum	20 lbs/hr
Type	Dry volumetric feed
Control	Manual-automatic variable delivery

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Miscellaneous	
Level indicators and vibrators on lime hopper; level indicators to control operation of storage bin feed gate; feeder discharge to be paced by pH measurement in oxidation and Densator units	
Lime slurry	
Mix tank	
Capacity	60 gallons
Construction	Steel
Miscellaneous	
Float control tied in with slurry water feed; lime feeder, and lime slurry pumpl flash mixer	
Pump	
Type	Positive displacement, manual variable drive
Capacity	15 gpm
Construction	Caustic and acid resistant
Piping	Flexible hose, arrange for flushing with raw mine drainage
Discharge	Flash mixer, Densator unit primary reaction zone and influent sump
Miscellaneous	
Slurry water from treated mine drainage sump, or well; flow indicator on slurry water and lime slurry lines	
Quicklime Storage and Feed	
Type of quicklime	High calcium and dolomitic pebble, to be fed separately
Storage bin	
Capacity	25 tons
Miscellaneous	
Tank truck delivery (17-20); blower hose; screw discharge to lime hopper; level indicators; dust collectors; circulating fan; vibrators	

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Lime feeder	
Feed rate	
Maximum	350 lbs/hr
Minimum	20 lbs/hr
Type	Dry volumetric feed by scres
Control	Manual-automatic variable delivery
Miscellaneous	
Feeder discharge to be paced by pH measurement in oxidation and Densator units	
Lime slaker	
Capacity	300 lbs/hr
Mix ratio	4 water:1 lime, by weight
Grit removal	Manual
Pump	
Type	Positive displacement, manual variable drive
Capacity	15 gpm
Construction	Caustic and acid resistant
Piping	Flexible hose, arrange for flushing with raw mine drainage
Discharge	Flash mixer, Densator unit primary reaction zone and influent sump
Miscellaneous	
Dust collector and vapor condenser; mixers; slurry water from treated mine drainage sump, or well; flow indicator on slurry water and lime slurry lines; float control in slaker tied in with slurry water feed, quicklime feeder, and quicklime slurry pump	
Sodium Hydroxide Storage and Feed	
Storage for 20-50 gallon drums in control building, additional storage in garage	

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Pump	
Type	Positive displacement, manual variable drive
Suction	Sodium hydroxide drum
Capacity	0.2 to 1.0 gpm of 50% NaOH, specific gravity of 1.54
Construction	Caustic Resistant
Piping	PVC
Discharge	Flash mixer, Densator unit primary reaction zone and influent sump
Sodium Carbonate Storage and Feed	
Storage hopper	
Capacity	1,500 lbs.
Miscellaneous	
Hand loading from ground floow with 100 pound bags; level indicators; dust collector, vibrators, storage for 50-100 pounds bags in control building, additional storage in garage	
Feeder	
Feed rate	
Maximum	50 lbs/hr
Minimum	20 lbs/hr
Type	Dry volumetric feed
Control	Manual variable delivery
Sodium carbonate slurry	
Mix tank	
Capacity	60 gallons
Construction	Steel
Miscellaneous	
Float control tied in with slurry water feed, sodium carbonate feeder, and sodium carbonate slurry pump; flash mixer	
Pump	
Type	Centrifugal, manual variable drive
Capacity	15 gpm
Construction	Caustic and acid resistant

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Piping	PVC
Discharge	Flash mixer, Densator unit primary reaction zone and influent sump
Miscellaneous	
Slurry water from treated mine drainage sump, or well; flow indicator on slurry water and sodium carbonate slurry lines	
Polyelectrolyte Storage and Feed	
Storage for 6-100 pound bags in control building	
Polyelectrolyte slurry	
Mix tank and flash mixer	
Volume	100 gallons
Construction	Steel
Slurry pump	
Type	Positive displacement, manual variable drive
Suction	Flooded
Capacity	0.06 to 2 gpm of 3% polyelectrolyte slurry
Construction	Standard
Piping	PVC
Discharge	Mixing tee
Dilution water	
Volume	5 gpm
Miscellaneous	
Supply from plant water pump, or well; discharge to settling lagoon, Densator unit primary reaction zone and influent sump, and sludge conditioning tank	
Plant Water Pumps and Sump	
Sump	
Capacity	1,000 gallons
Construction	Concrete
Miscellaneous	
Sump to receive discharge from settling lagoon and Densator unit; overflow to Tyler Run; continuous pH, temperature, iron and turbidity recording; rate-of-flow control on holding lagoon discharge line to close at pre-set pH level	

Table 1 (continued). UNITS AND FACILITIES FOR COLLECTING AND TREATING MINE DRAINAGE DISCHARGES - PRELIMINARY BASIS OF DESIGN

Pumps	
Number	2 (1 on stand-by)
Type	Centrifugal, constant speed with by-pass
Suction	Flooded
Capacity	70 gpm
Construction	Standard
Piping	Cast iron
Discharge	Control building plant water piping
Sludge Drying Lagoon	
Construction	Earth placed in compacted layers, mechanical lines at inlets, drainage provided by 1 foot of sand and 3 feet of culm underlying lagoon
Freeboard	2 feet
Side slopes	2:1
Compartments	
Number	3
Dimensions (L x W x SWD)	50 x 25 x 3 feet
Volume/compartment	34,000 gallons
Supernatant overflow	Tyler Run
Piping	PVC
Miscellaneous	
Locate top of sand above 1205 elevation; provide access road around lagoon; cover 1 compartment (wooden trusses and plastic covering - opening on 4 sides); lagoon to receive sludge from settling lagoon and Densator unit, and filtrate from vacuum filter; drainage channel to Tyler Run for underflow	

Table 2. SPECIFICATIONS FOR SURFACE AERATORS[a]
(Lightnin Model LAR-60)

Impeller Diameter	59" (4-blade - 316 S.S.)
Impeller Speed	19-56 RPM (variable)
Impeller Motor Drive	7-1/2 H.P. - 1750/583 RPM
Oxygen Transfer Rate	30 lbs O_2 @ STP/hr
	4.0 lbs O_2/hr/H.P.
Pumping Capacity	20,000 GPM @ max. speed
Impeller Elevation	Adjustable ± 6" in 1" increments from normal position

[a] The specification oxygen transfer rate and liquid pumping capacity are valid for the impeller position as shown in Figure 13 where the top of the impeller blade is at the liquid level and with the indicated tank geometry. If the impeller immersion is increased, the pumping rate would increase but the oxygen transfer rate would decrease. If the impeller is raised, both the pumping capacity and the oxygen transfer rate would decrease. Since each oxidation tank has a capacity of approximately 10,000 gallons, the nominal pumping rate of the aerators in normal position produced a complete tank turn-over every one-half minute.

Table 3. POWER CONSUMPTION BY VARIOUS UNIT OPERATIONS

Designation	Power rating, hp	Current rating amperes	Voltage	Measured power consumption KWHr/Hr			Cost[a] (cents/hr.)		
				average	high	low	average	high	low
Pumps:									
Water Collection									
Proctor No. 1	15	19.0	440	3.98	9.67	0.75	6.8	16.4	1.3
Proctor No. 2	25	31.0	440	6.59	17.68	0.94	11.2	30.1	1.6
Tyler Run	25	31.0	440	6.68	9.79	1.58	11.4	16.6	2.7
Bennett's Branch	30	36.6	440	6.09	17.60	0.12	10.4	29.9	0.2
Process									
Vacuseal Pump	10	13.5	440						
Densator Feed	10	16.5	440	1.16	2.93	0.11	2.0	5.0	0.2
Lime Slurry Circulation	10	13.5	440	3.60	8.73	0.17	6.1	14.8	0.3
Precoat Feed	2	4.2	440	0.29	0.33	0.25	0.5	0.6	0.4
Filter Feed	2	3.6	440	0.20	0.60	0.01	0.3	1.0	0.02
Process Water (2 motors)[b]	7.5	9.8	440	2.46	5.23	0.13	4.2	8.9	0.2
Compressor - Housewater	0.75	1.55	440	0.10*			0.2		
Compressor - Autovalve	1.0	13.8	115	1.00*			1.7		
Sample Pump No. 1	0.75	9.0	115	0.70*			0.1		
Sample Pump No. 2	0.75	9.0	115	0.70*			0.1		
Drain Sump	0.33	1.5	440	0.50*			0.1		
Sludge Sump Drain	0.33	5.4	115	0.40			0.1		
Chlorinator	--	1.2	115	0.01*			0.01		
Unit Operations:									
Densator									
Mixers	1.5	3.3	440	0.38	0.57	0.10	0.6	1.0	0.2
Sludge Scraper	0.5	1.4	440	0.16	0.33	0.03	0.3	0.6	0.05
Recirculation	1.5	3.3	440	0.34	2.40	0.02	0.6	4.1	0.03
Total	3.5	8.0		0.88	3.30	0.15	1.5	5.6	0.3

Table 3 (continued). POWER CONSUMPTION BY VARIOUS UNIT OPERATIONS

Designation	Power rating, hp	Current rating amperes	Voltage	Measured power consumption KWHr/Hr			Cost[a] (cents/hr.)		
				average	high	low	average	high	low
Vacuum Filter									
Vacuum Pump	30	36.5	440	10.70	17.40	1.88	18.2	29.6	3.2
Agitator	1.5	2.4	440	0.22	0.40	0.07	0.4	0.7	0.1
Knife Retractor	1	1.8	440						
Drum Drive	1	2.8	440	0.22	0.40	0.10	0.4	0.7	0.2
Filtrate Pump	3	4.6	440	1.07	1.40	0.32	1.8	2.4	0.5
Sludge Conveyor	2	3.6	440	0.19	0.42	0.08	0.3	0.7	0.1
Conditioning Tank Mixer	0.33	4.0	115						
Total	38.83	55.7		17.07	27.70	3.29	29.0	47.1	5.6
Chemical Oxidation									
Surface Aerator No. 1	7.5	10.5	440	1.82	11.11	0.42	3.1	18.9	0.7
Surface Aerator No. 2	7.5	10.5	440	1.36	4.40	0.13	2.3	7.5	0.2
Reactors									
Limestone Rotary Reactor	20.0	25.6	440	7.49	20.20	0.10	12.7	34.3	0.2
Limestone Reactor Feeder	--	5.0	220	0.8*			1.4		
Flash Mixer Agitator No. 1	0.33	5.4	115	0.4*			0.5		
Flash Mixer Agitator No. 2	0.33	5.4	115	0.4*			0.5		
Reagent Control									
Hydrated Lime									
Lime Conveyor No. 1	1	2.5	440	0.05	0.07	0.04	0.09	0.1	0.07
Lime Conveyor No. 2	1	1.8	440	0.04	0.05	0.04	0.07	0.09	0.07
Mixer	0.5	1.1	440	0.17*			0.09		
Helix	0.25		115	0.30*			0.5		
Air Blower	5	7.1	440	0.05	0.07	0.04	0.09	0.1	0.07
Vibrator	--	2.0	115	0.16*			0.3		
Recirculation Fan	1	1.82	440	0.41*			0.7		
Total	8.75	16.3		1.23*			0.9		

Table 3 (continued). POWER CONSUMPTION BY VARIOUS UNIT OPERATIONS

Designation	Power rating, hp	Current rating amperes	Voltage	Measured power consumption KWHr/Hr average	high	low	Cost[a] (cents/hr.) average	high	low
Pebble Lime									
Vibrator Conveyor - 2		1.75	115	0.13*					
Lime Conveyor No. 3	1	1.8	440	0.03	0.03	0.02	0.05	0.05	0.03
Mixer	1.5	2.3	440	0.51	0.71	0.13	0.9	1.2	0.2
Helix	0.25		115	0.30*			0.5		
Grit Removal	0.25	5.2	115	0.40*			0.7		
Recirculation Fan	1	1.82	440	0.56*			1.0		
Total	4.0	12.87		1.93*			3.3		
Liquid Reagents									
Distribution Pump	0.33	5.6	115	0.43*			0.7		
Sludge Control									
Transfer Pump (Settling Lagoon)	3	5.1	440	0.15	0.29	0.04	0.3	0.5	0.07
Sludge Sump Mixer	0.5	5.5	220	0.85*			1.4		
Precoat Feeder	0.25	4.4	115	0.34*			0.6		
Precoat Agitator	0.25	3.2	220	0.48*			0.8		
Precoat Vibrator	--	2.0	115	0.15*			0.3		
Flocculant Preparation									
Mixer	0.33	5.6	115	0.43*			0.7		
Distribution Pump	0.5	1.1	440	0.34*			0.6		

[a] Based on 1.7¢ per KWHr. Actual rate schedule varies with demand and consumption.

[b] Ratings are for the individual motor, consumptions for the system.

*Estimated.

Table 4. DIMENSIONS OF SETTLING LAGOON COMPARTMENTS

			Volume				
	Top Dimensions	Area	Sludge Storage		Total		Total Cumulative
	ft.	sq. ft.	cu. ft.	gallons	cu. ft.	gallons	gallons
Influent Section	52W x 65L	3,380	5,400	40,395	19,240	143,925	143,925
Middle Section	52W x 50L	2,600	5,000	37,403	16,500	123,429	267,354
Effluent Section	70W x 54L	3,780	--	--	23,516	175,927	443,281

Pond Depth - 5 ft. minimum - 8 ft. maximum (at center of lagoon)

Average Depth = approx. 6 ft.

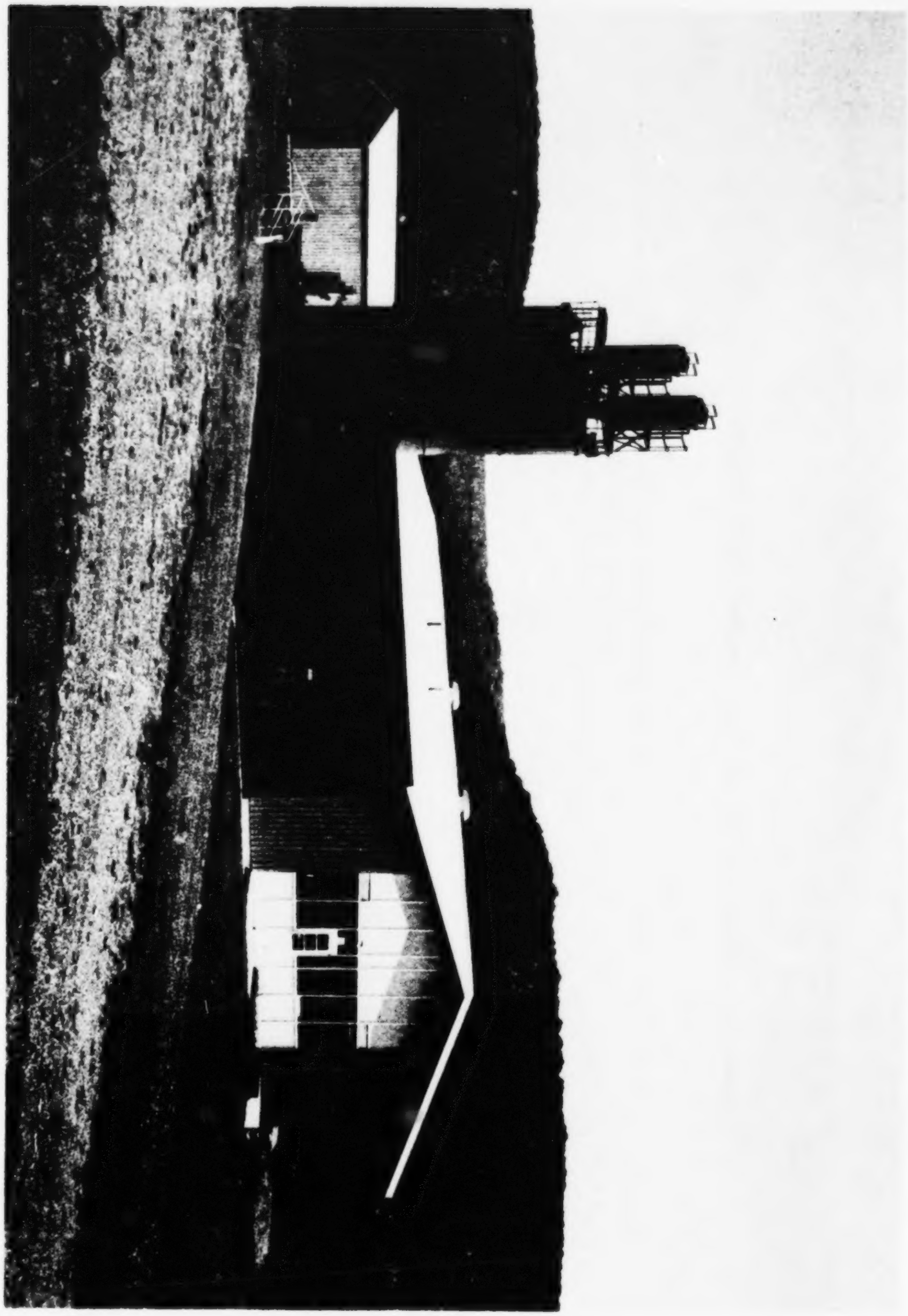


Figure 1. Photograph of Control Building

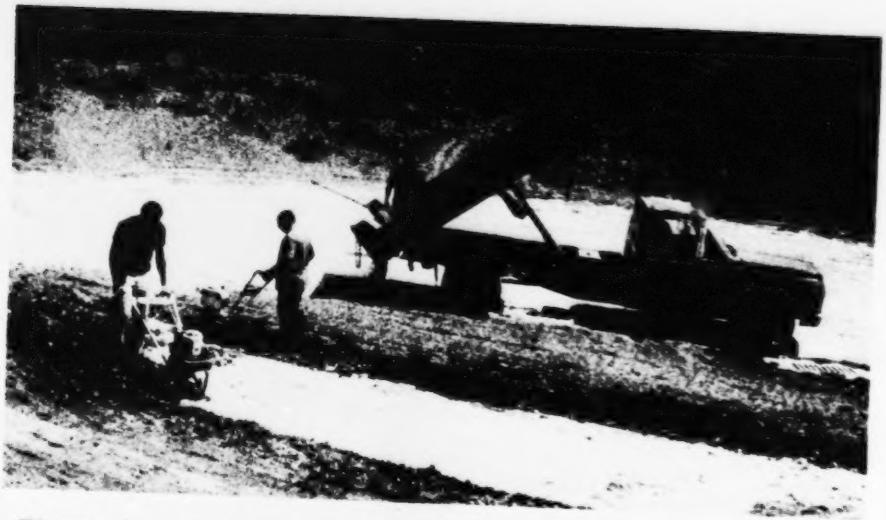


Figure 2. Application of Bentonite Clay to Holding Lagoon

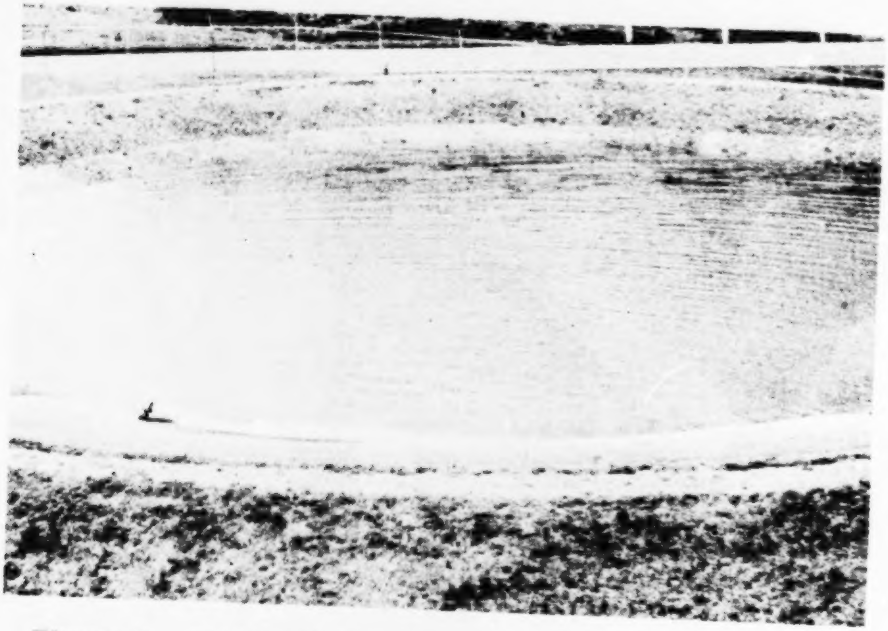


Figure 3. Filling Holding Lagoon

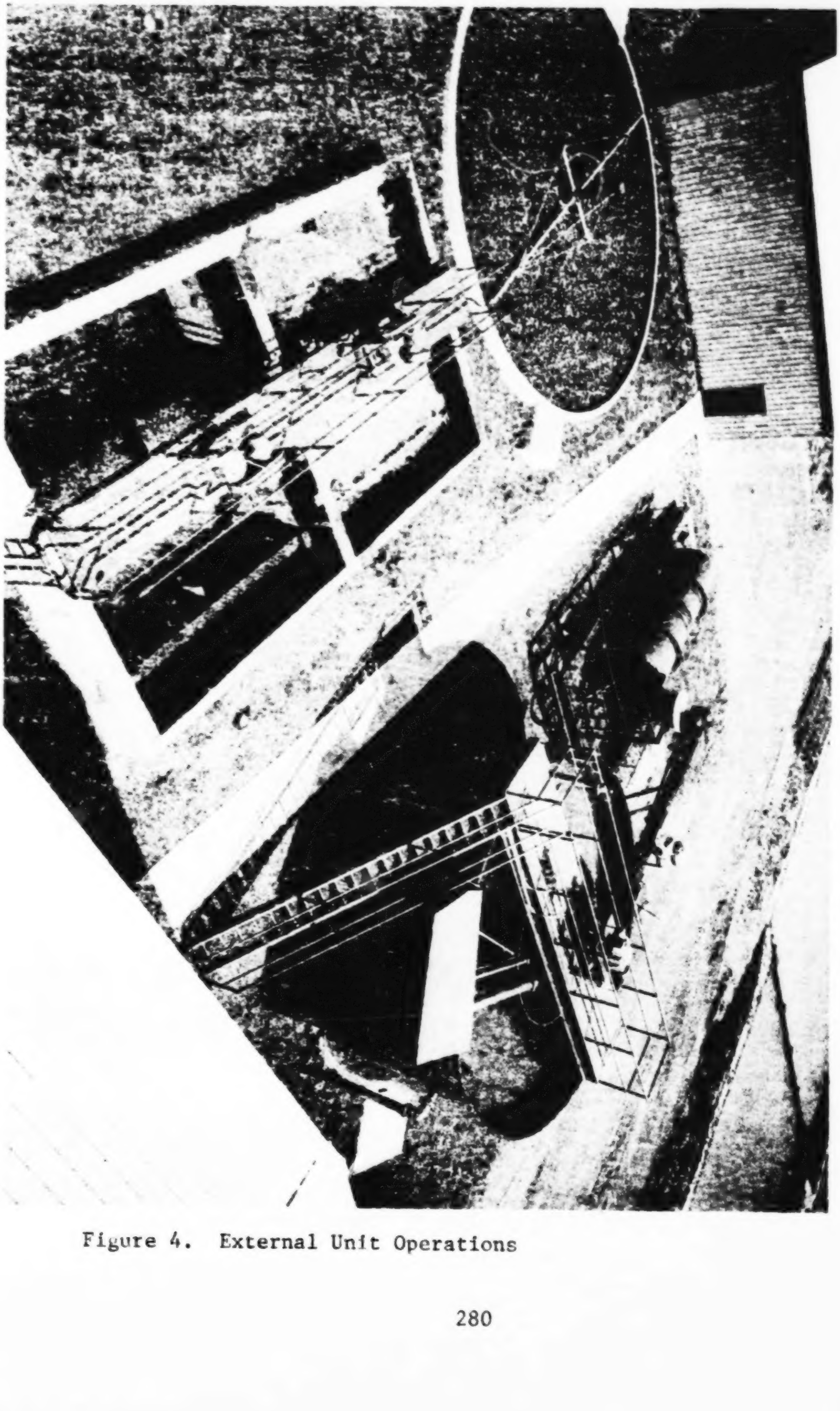


Figure 4. External Unit Operations

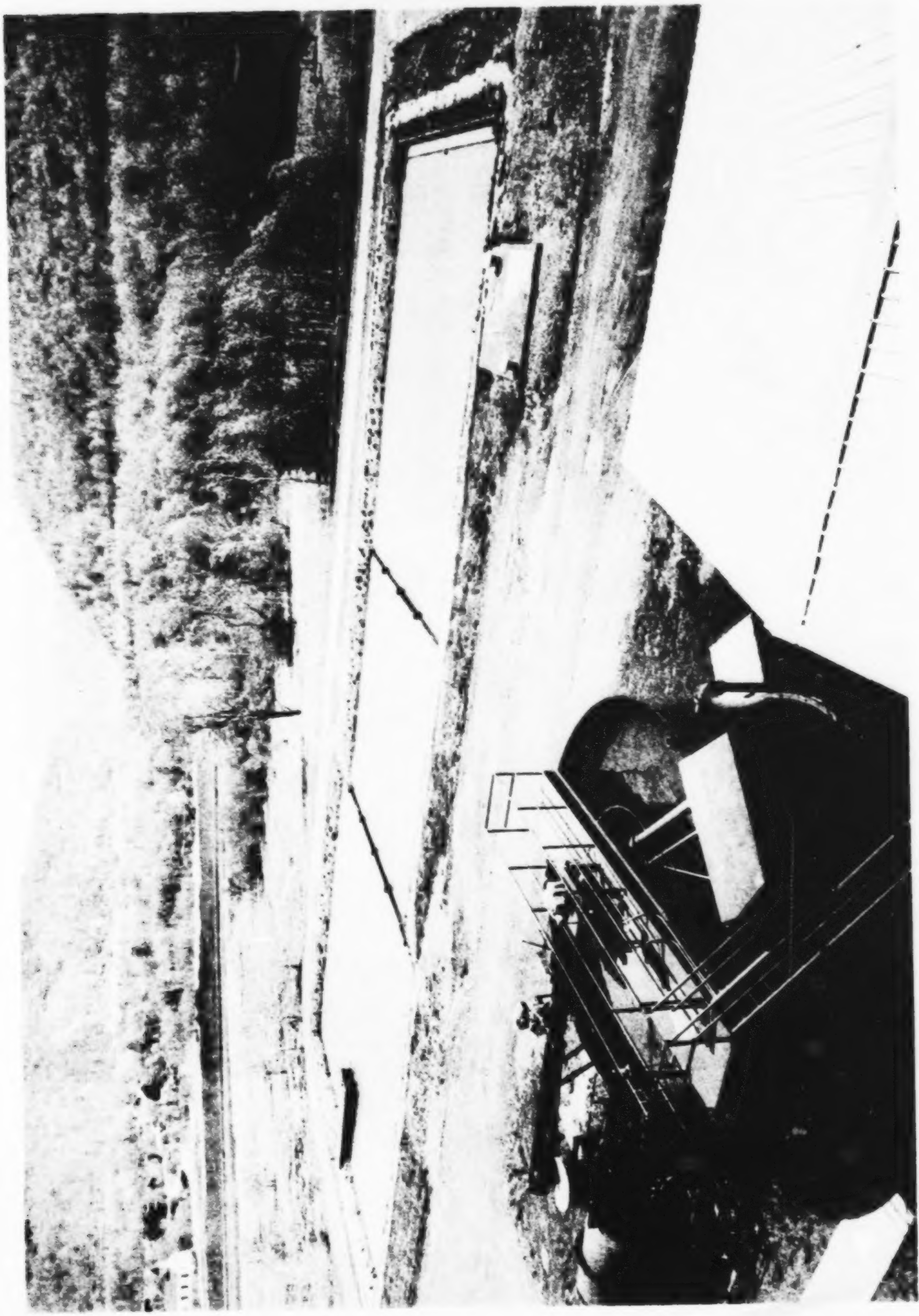


Figure 5. The Settling Lagoon

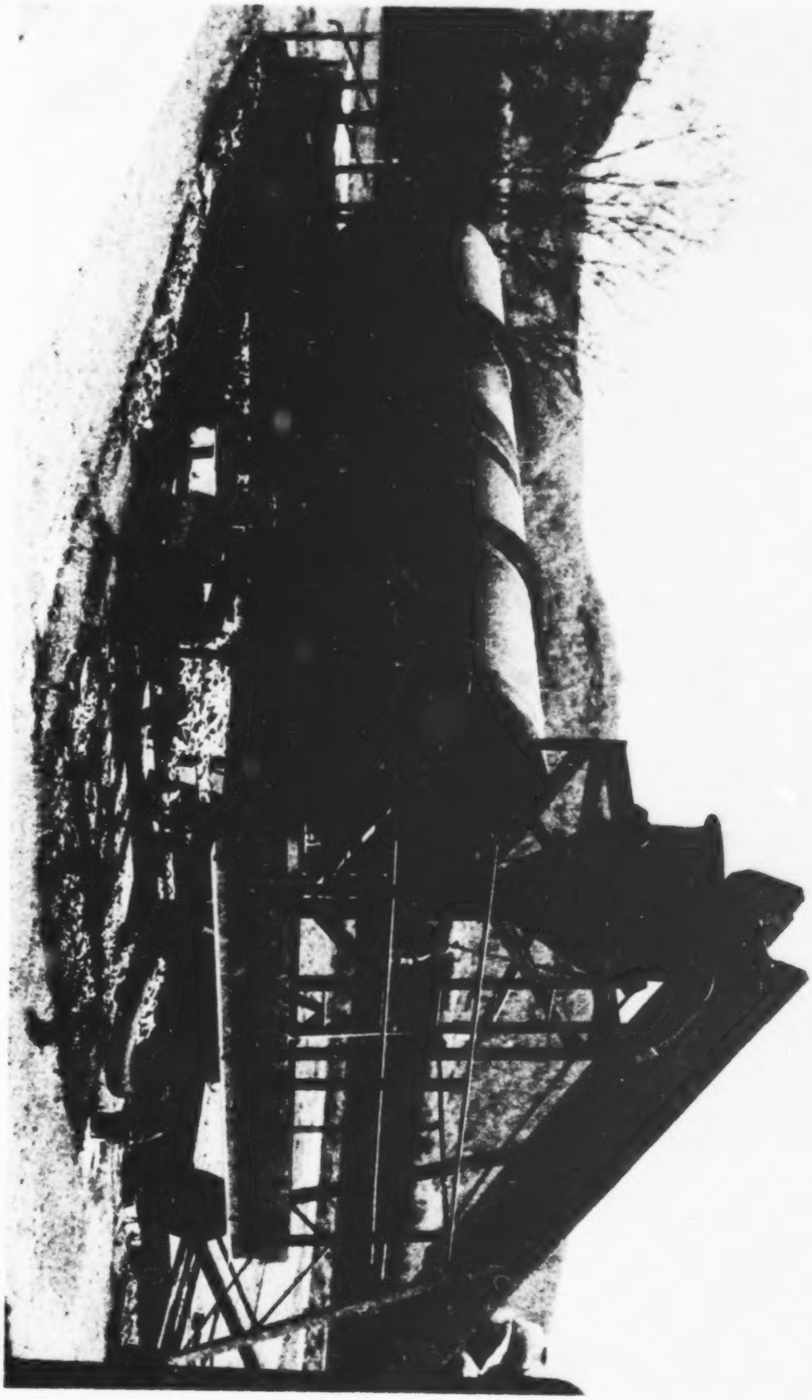


Figure 6. Limestone Reactor Showing Effluent Detail

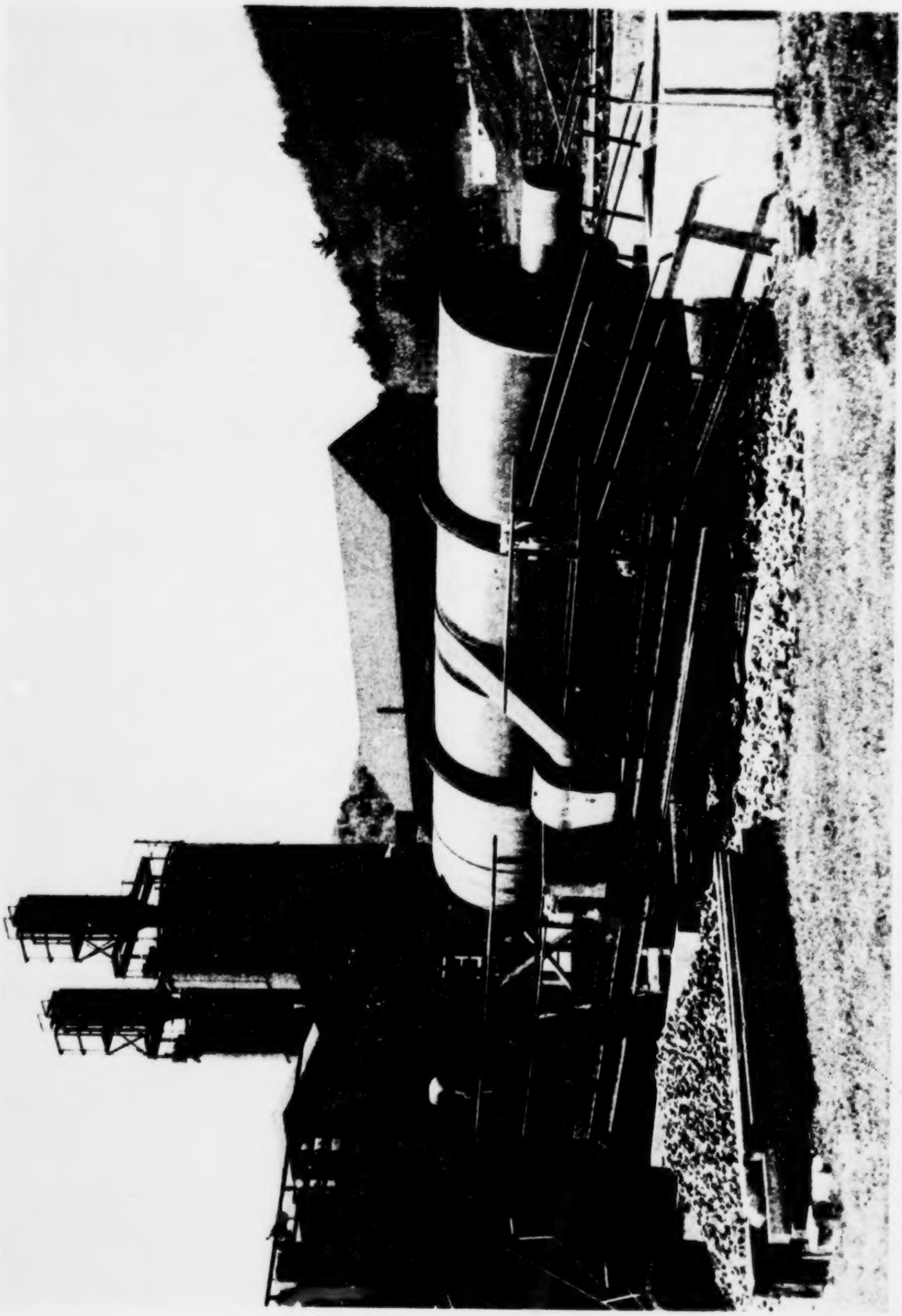


Figure 7. Limestone Reactor Showing Effluent Detail and Drive Mechanism

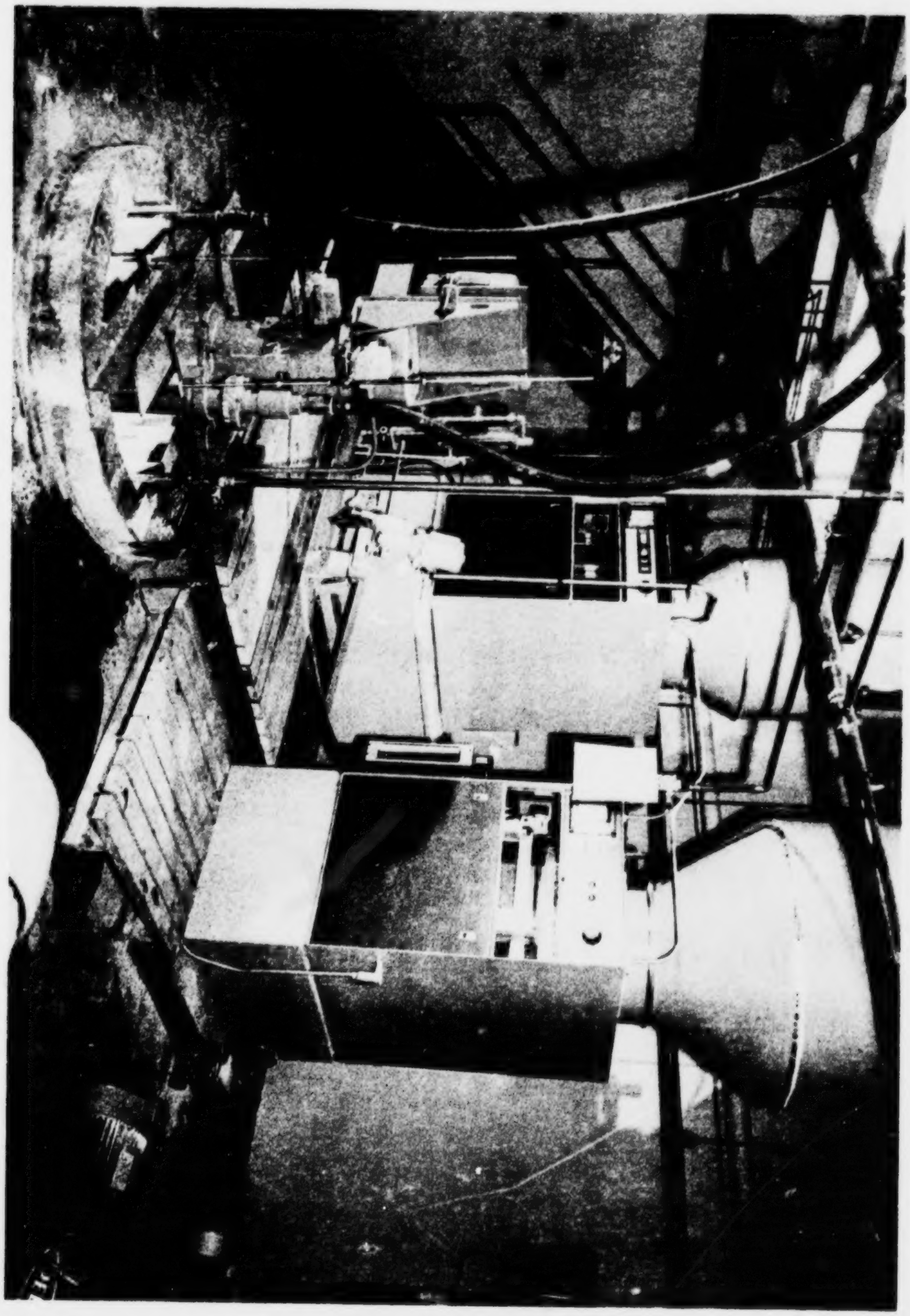


Figure 8. Alkali Slurry Preparation Equipment

Reproduced from best available copy.

Figure 9. Internal piping and unit operations

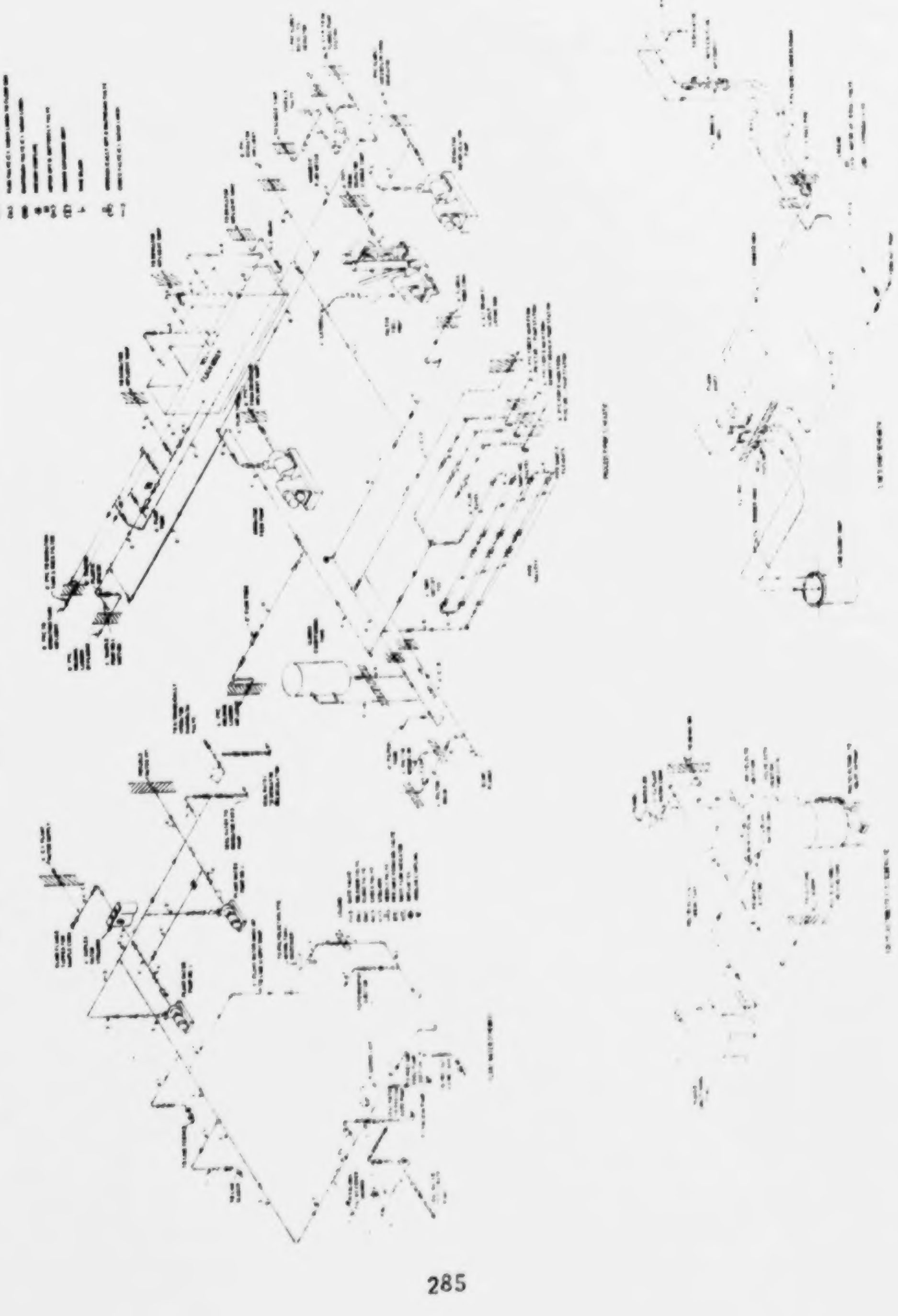
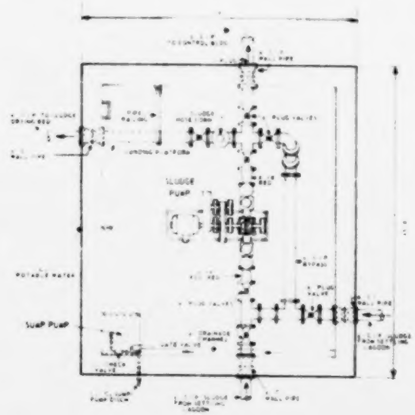
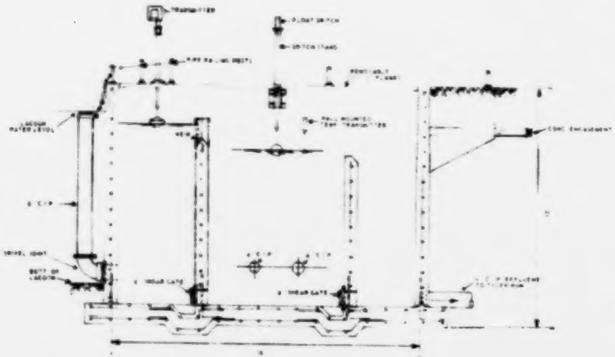


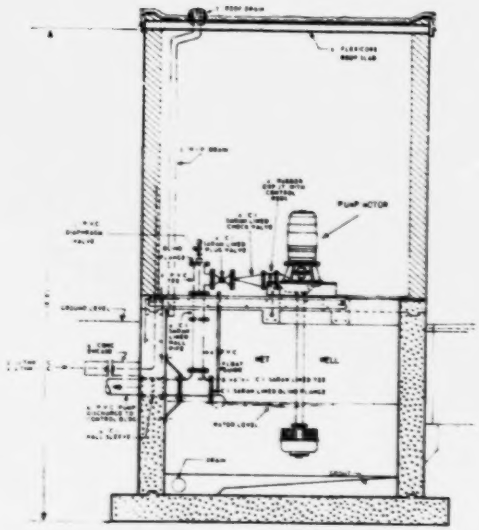
Figure 10. Sump designs


SLUDGE
PUMP
SUMP PUMP
POTABLE WATER
WALL PIPE
PLUG VALVES
BYPASS
PLUG VALVE
GATE VALVE
CHECK VALVE
DRAINAGE CHANNEL
TO CONTROL BLDG
TO SLUDGE DRYING BED
FROM SETTLING LAGOON

SLUDGE PUMPING STATION
Scale 3/8" - 1'0"

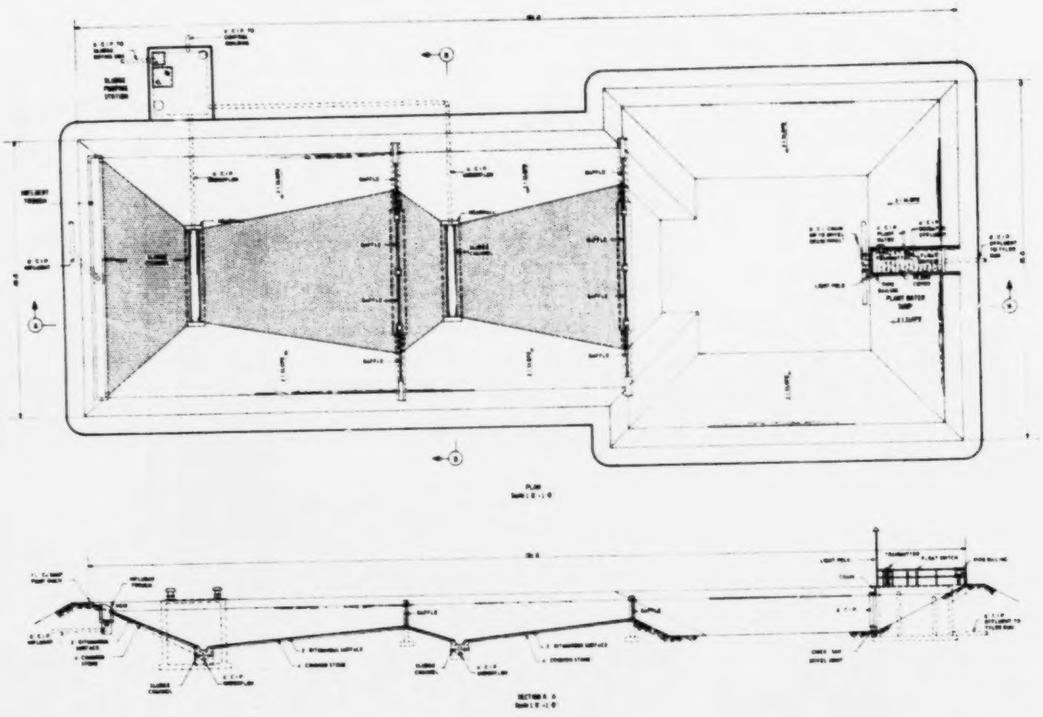

TRANSMITTER
FLOAT SWITCH
SWITCH STAND
PIPE RAILING (BEST)
REMOVABLE PLANK
WALL MOUNTED TEMP TRANSMITTER
LAGOON WATER LEVEL
WEIR
SWIVEL JOINT
BOTT OF LAGOON
SHEAR GATE
SHEAR GATE
EFFLUENT TO FILTRATION
CONC ENCASEMENT

SETTLING LAGOON EFFLUENT STRUCTURE
Scale 3/8" - 1'0"


ROOF DRAIN
FLEXICORE ROOF SLAB
DRAIN
RUBBER COUPLING WITH CONTROL ROD
PUMP MOTOR
SARAN LINED CHECK VALVE
SARAN LINED PLUG VALVE
PVC DIAPHRAGM VALVE
BLIND FLANGE
PVC TEE
GROUND LEVEL
CONC ENCASE
SARAN LINED WALL PIPE
PVC
FLOAT GUARD
WET
WELL
SARAN LINED TEE
SARAN LINED BLIND FLANGE
PVC PUMP DISCHARGE TO CONTROL BLDG
WALL SLEEVE
WATER LEVEL
DRAIN
GROUT

PROCTOR NO. 1 PUMPING STATION
Scale ½" - 1'0"

Figure 11. Settling lagoon design


LIGHT POLE
PLANT WATER SUMP

Reproduced from best available copy.

Figure 12. Sludge drying basin design

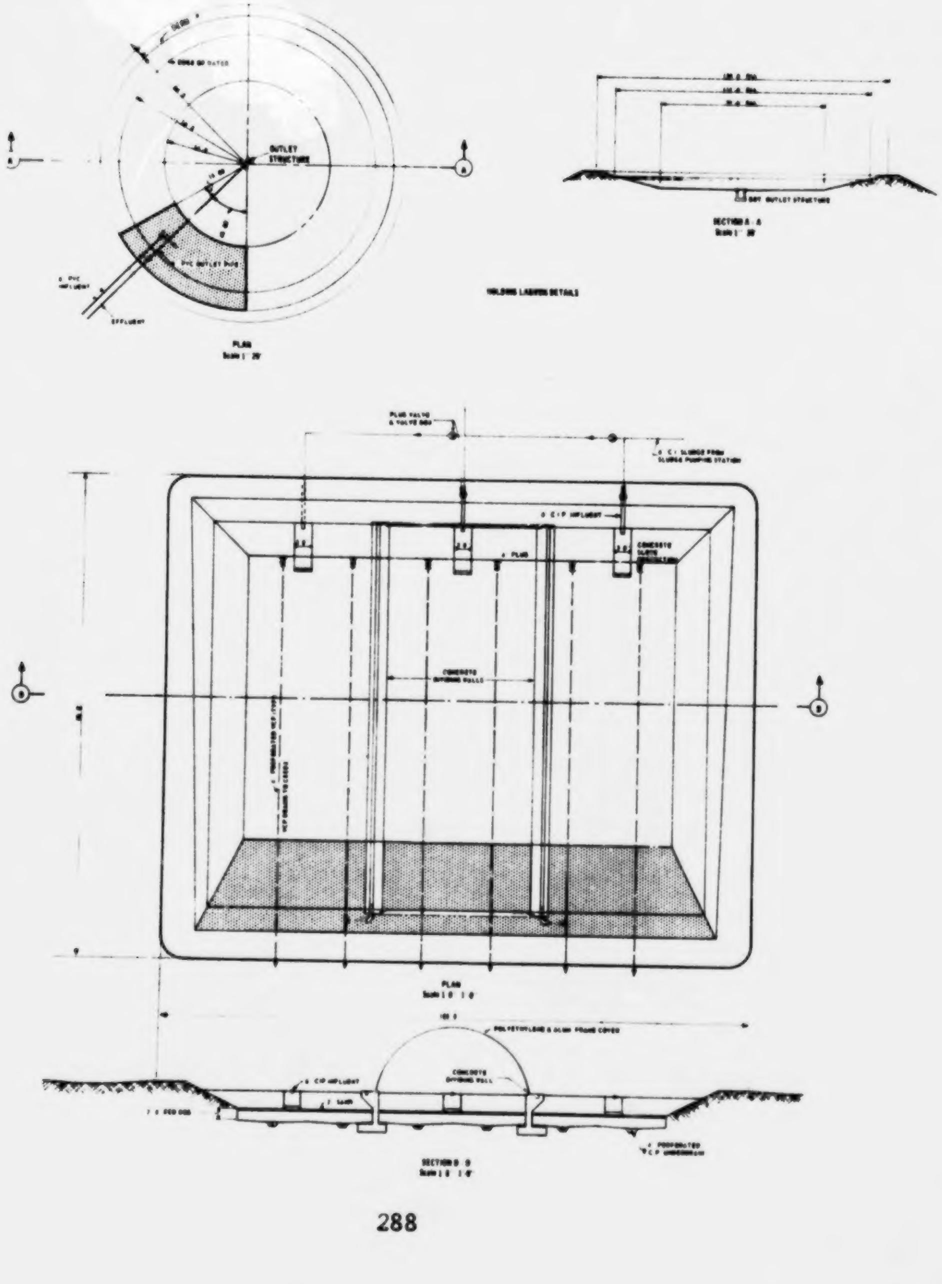
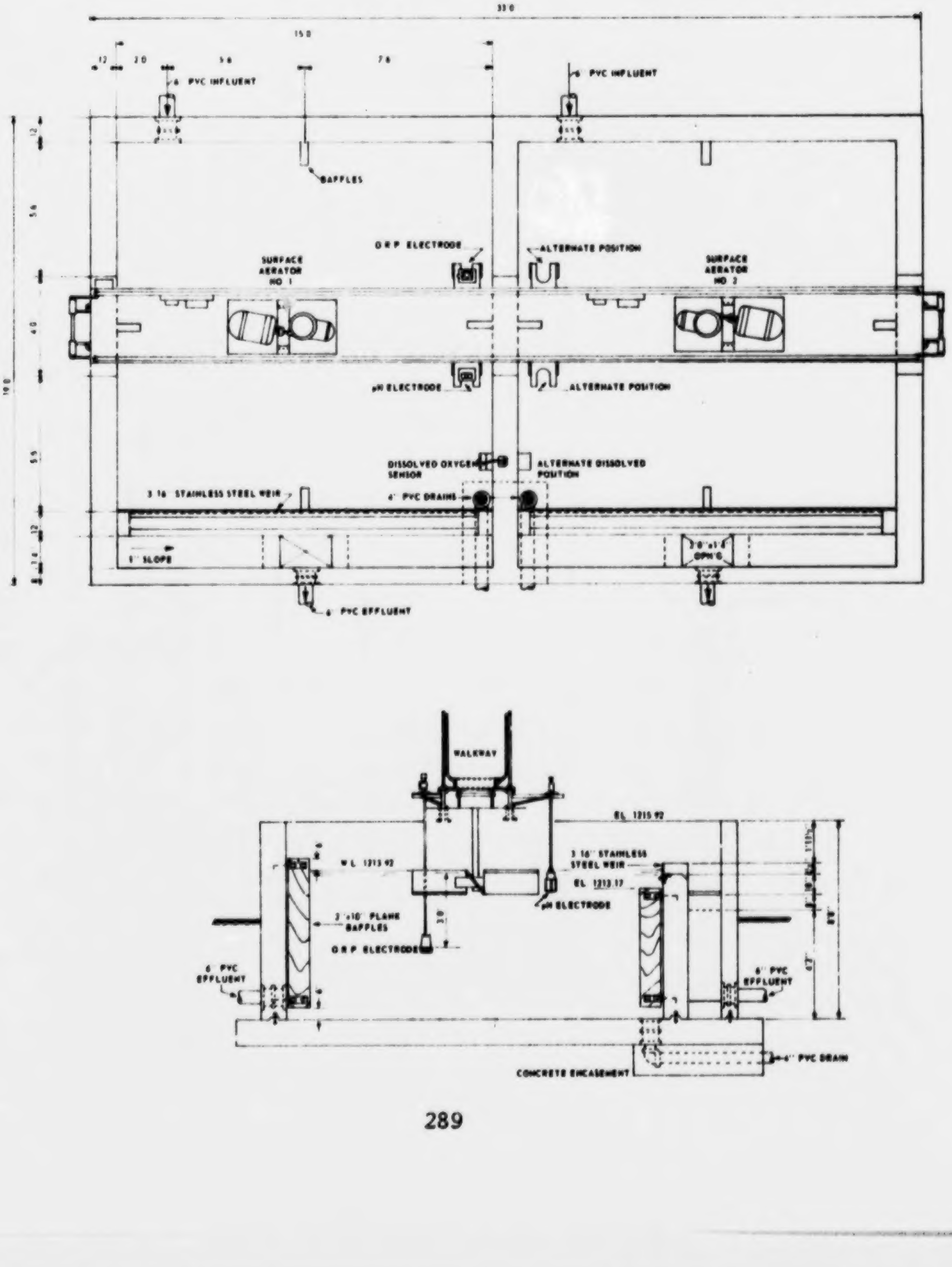

OUTLET STRUCTURE
A
A
PVC OUTLET PIPE
PVC INFLUENT
EFFLUENT
PLAN
SECTION A - A
OUTLET STRUCTURE
HOLDING LAGOON DETAILS
PLUG VALVE & VALVE BOX
SLUDGE FROM SLUDGE PUMPING STATION
C I P INFLUENT
CONCRETE SLUTE
PLUG
CONCRETE DRYING WALLS
B
B
PERFORATED VCP
PLAN
POLYETHYLENE & ALUM FRAME COVER
CIP INFLUENT
SAND
CONCRETE DRYING WALL
PERFORATED VCP UNDERDRAIN
SECTION B - B

Figure 13. Oxidation tanks with surface aerator



33 0
15 0
12
2 0
5 6
7 6
6" PVC INFLUENT
6" PVC INFLUENT
BAFFLES
O R P ELECTRODE
ALTERNATE POSITION
SURFACE AERATOR NO 1
SURFACE AERATOR NO 2
pH ELECTRODE
ALTERNATE POSITION
DISSOLVED OXYGEN SENSOR
ALTERNATE DISSOLVED POSITION
3/16" STAINLESS STEEL WEIR
4" PVC DRAINS
1" SLOPE
2'0"x1'4" OPN'G
6" PVC EFFLUENT
WALKWAY
EL 1215.92
W L 1213.92
3/16" STAINLESS STEEL WEIR
EL 1213.17
pH ELECTRODE
2"x10" PLANK BAFFLES
O R P ELECTRODE
3 0
6" PVC EFFLUENT
6" PVC EFFLUENT
CONCRETE ENCASEMENT
4" PVC DRAIN

Figure 14. Infilco densator design

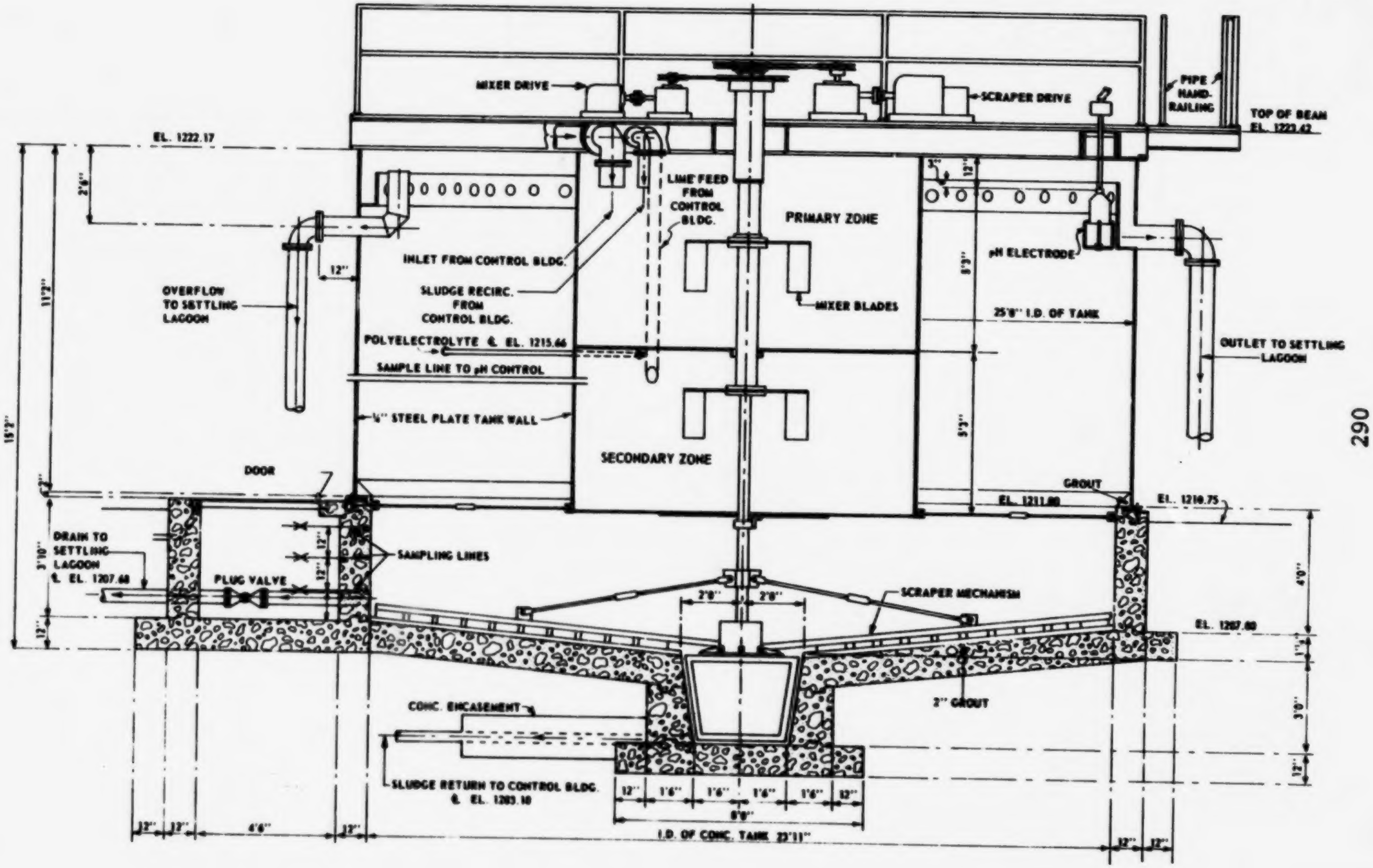

MIXER DRIVE
SCRAPER DRIVE
PIPE HAND-RAILING
TOP OF BEAM EL. 1223.42
EL. 1222.17
2'6"
11'2"
15'2"
12"
OVERFLOW TO SETTLING LAGOON
INLET FROM CONTROL BLDG.
SLUDGE RECIRC. FROM CONTROL BLDG.
LIME FEED FROM CONTROL BLDG.
POLYELECTROLYTE ℄ EL. 1215.66
SAMPLE LINE TO pH CONTROL
¼" STEEL PLATE TANK WALL
PRIMARY ZONE
MIXER BLADES
SECONDARY ZONE
3"
12"
pH ELECTRODE
8'3"
25'8" I.D. OF TANK
5'3"
OUTLET TO SETTLING LAGOON
DOOR
GROUT
EL. 1211.80
EL. 1210.75
DRAIN TO SETTLING LAGOON ℄ EL. 1207.68
3'10"
PLUG VALVE
SAMPLING LINES
12"
12"
12"
2'8"
2'8"
SCRAPER MECHANISM
EL. 1207.80
4'0"
1'½"
3'0"
12"
2" GROUT
CONC. ENCASEMENT
SLUDGE RETURN TO CONTROL BLDG. ℄ EL. 1203.10
12"
1'6"
1'6"
1'6"
1'6"
12"
8'0"
12"
12"
4'6"
12"
I.D. OF CONC. TANK 23'11"
12"
12"

Figure 15. Surface biochemical oxidation reactor

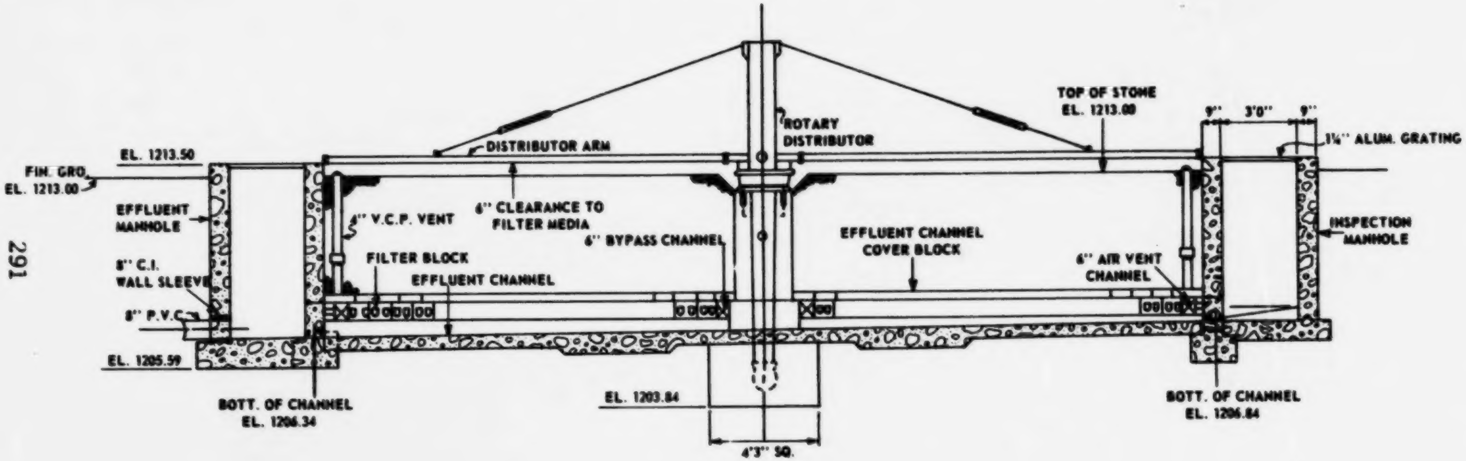
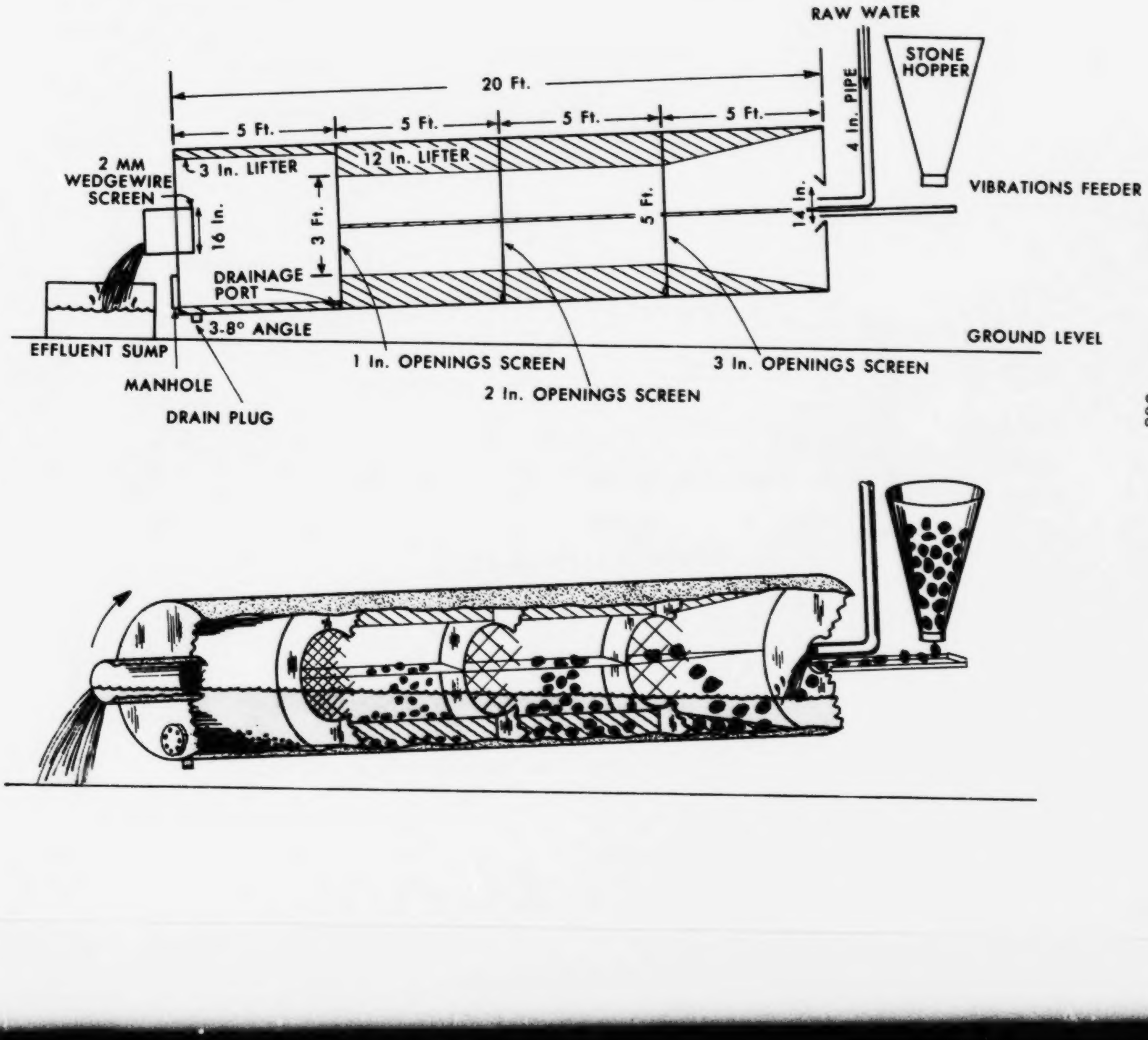

TOP OF STONE
EL. 1213.00
ROTARY
DISTRIBUTOR
9''
3'0''
9''
1¼'' ALUM. GRATING
DISTRIBUTOR ARM
EL. 1213.50
FIN. GRD.
EL. 1213.00
EFFLUENT
MANHOLE
4'' V.C.P. VENT
6'' CLEARANCE TO
FILTER MEDIA
6'' BYPASS CHANNEL
EFFLUENT CHANNEL
COVER BLOCK
INSPECTION
MANHOLE
8'' C.I.
WALL SLEEVE
FILTER BLOCK
EFFLUENT CHANNEL
6'' AIR VENT
CHANNEL
8'' P.V.C.
EL. 1205.59
BOTT. OF CHANNEL
EL. 1206.34
EL. 1203.84
BOTT. OF CHANNEL
EL. 1206.84
4'3'' SQ.

Figure 16. Schematic of Limestone Reactor


RAW WATER
STONE HOPPER
20 Ft.
5 Ft.
5 Ft.
5 Ft.
5 Ft.
4 In. PIPE
2 MM WEDGEWIRE SCREEN
3 In. LIFTER
12 In. LIFTER
VIBRATIONS FEEDER
16 In.
3 Ft.
5 Ft.
14 In.
DRAINAGE PORT
3-8° ANGLE
GROUND LEVEL
EFFLUENT SUMP
MANHOLE
DRAIN PLUG
1 In. OPENINGS SCREEN
2 In. OPENINGS SCREEN
3 In. OPENINGS SCREEN

APPENDIX C

DETAILED PLANT DATA

Table 1. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME PROCESS CONDITIONS (PROCESS 10)

Water source	Average flow, gpm	Run duration hrs	Water treated gal	Test No.
Tyler Run	182	47.5	520,912	10-70
	247	67.5	1,001,060	9-70
Bennett's Branch	153	48.0	441,400	3-70
	240	48.0	693,600	2-70
	317	90.0	1,716,650	1-70
Proctor No. 1	112	18.0	120,636	8-70
	182	24.0	263,200	4-70
	251	60.0	904,400	5-70
	298	48.0	859,570	6-70
Proctor No. 1 and Bennett's Branch	342 (290+52)	48.0	984,960	7-70

Process Flow:

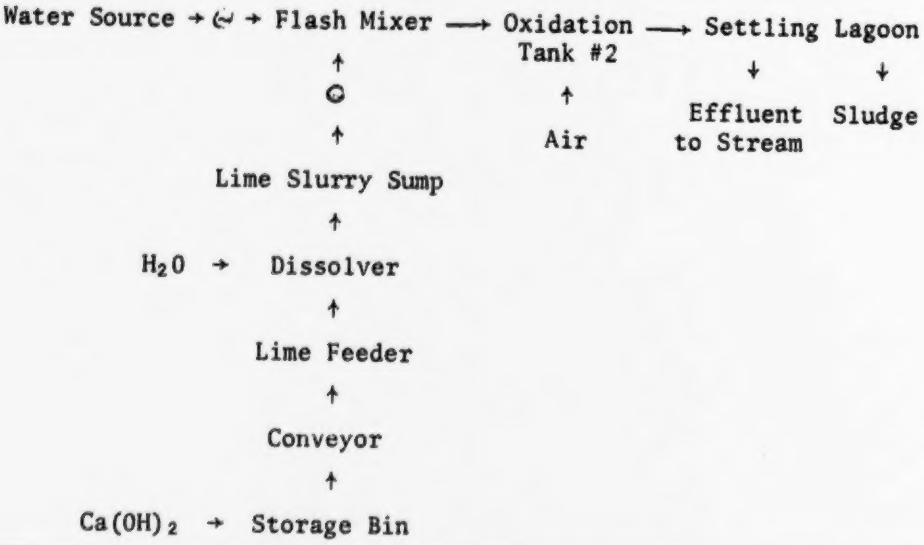

Water Source
Flash Mixer
Oxidation Tank #2
Settling Lagoon
Effluent to Stream
Sludge
Air
Lime Slurry Sump
H_2O
Dissolver
Lime Feeder
Conveyor
$Ca(OH)_2$
Storage Bin

Table 2. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME - RAW WATER ANALYSES

Test No.	9-70	10-70	3-70	2-70	1-70	8-70	4-70	5-70	6-70	7-70
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 1 Bernetts Branch
Temp.°F, Av.	60	57	57	55	51	52	52	51	55	53
Range	56-65	56-61	55-60	51-58	46-57	51-54	50-53	48-53	52-60	52-55
pH, Av.	3.96	3.47	4.08	4.05	3.96	3.50	3.32	3.29	3.29	3.28
Range	3.40-6.98	3.45-3.49	4.03-4.12	3.91-4.33	3.67-4.73	3.42-3.54	3.29-3.42	3.25-3.33	3.26-3.31	
Acidity, Av.	89	90	74	68	101	446	511	513	599	384
mg/l Range	89-93	89-91	70-80	57-76	12-139	418-474	502-519	451-580	450-1030	196-492
Fe II, Av.	1.9	1.8	13	13	23	55	58	60	60	49
mg/l Range	1-3	1-2	11-15	6-17	1.7-34	54-56	57-59	58-63	58-62	26-66
Fe III, mg/l, Av.	≥0*	≥0*	6*	6*	≥0*	≥0*	≥0*	≥0*	≥0*	≥0*
Al, mg/l, Av.	9*	9*	26*	26*	17*	80*	84*	84*	82*	73*
Ca, mg/l, Av.	59*	59*	36*	36*	36*	96*	96*	96*	96*	80*
Mg, mg/l, Av.	64*	64*	40*	40*	40*	76*	76*	76*	76*	60*
SO_4,mg/l, Av.	180*	180*	179*	179*	179*	730*	730*	730*	730*	600*
Conductivity, Av.	925	932	560	554	772	1186*	1170*	1174*	1182*	1806
micromhos Range	856-946	900-1000	470-630	470-640	320-1010					1760-1850

* Values established from monthly analyses.

Table 3. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME - PROCESS AND REAGENT REQUIREMENTS

Test No.	9-70	10-70	3-70	2-70	1-70	8-70	4-70	5-70	6-70	7-70
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No 1 Bennetts Br.
Influent Fe II										
lbs.	15.5	9.1	48.6	70.0	332.7	55.3	126.8	453.8	451.0	435.4
lbs./hr	0.235	0.191	1.018	1.448	3.697	3.072	5.282	7.563	9.396	9.072
Fe II Reacted, lbs	7.3	4.8	39.9	41.8	164.7	51.9	120.0	422.0	427.0	402.6
Minimum Oxidation Rate mg/l/hr	0.9	1.1	10.8	7.2	11.5	51.9	54.5	55.9	59.6	49.0
Theoretical Reagent Required[1]										
lbs	585.8	303.3	214.7	310.1	1139.9	353.7	884.3	3050.2	3385.0	2486.6
lbs/hr	8.6	6.5	4.5	6.5	12.7	19.7	36.8	50.8	70.5	51.8
lbs/1,000 gal	0.56	0.59	0.49	0.45	0.66	2.93	3.36	3.37	3.94	2.52
Reagent Utilized										
lbs	563	340	182	170	1350	680	986	2820	3020	3100
lbs/hr	8.3	7.2	3.8	3.5	15.0	37.8	41.1	47.0	62.9	64.6
lbs/1,000 gal	0.56	0.65	0.41	0.25	0.79	5.64	3.75	3.12	3.51	3.15
Alkali Consumption Utilized/Theoretical %	96.1	110:4	84.8	54.8	118.4	192.3	111.5	92.5	89.2	124.7
Theor. + Effl. Alk. Utilized %	107	103	67	156	71	55	82	103	111	85
Retention Time (minutes)										
Flash Mixer	6.0	8.1	9.7	6.2	4.7	13.2	8.1	5.9	5.0	4.3
Oxidation Tanks	40.2	54.5	64.8	41.4	31.3	88.6	54.5	39.6	33.3	29.0

[1] Based on 93.98% $Ca(OH)_2$

Table 3 (continued). TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME - PROCESS AND REAGENT REQUIREMENTS

Test No.	9-70	10-70	3-70	2-70	1-70	8-70	4-70	5-70	6-70	7-70
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 1 Bennetts Br.
Theoretical Oxygen Requirement										
lbs	2.2	1.3	7.0	10.0	47.7	7.9	18.2	65.0	64.6	62.4
lbs/hr	0.034	0.027	0.146	0.207	0.530	0.440	0.757	1.083	1.346	1.299
lbs/1,000 gal	0.0022	0.0025	0.0159	0.0144	0.0278	0.0655	0.0691	0.0719	0.0752	0.0634
ft^3 (std)	25.2	14.4	78.3	111.5	535.3	88.9	203.9	729.3	725.1	699.8
ft^3/hr (std)	0.38	0.30	1.64	2.32	5.95	4.94	8.50	12.16	15.11	14.58
Oxygen Utilized as Supplied by Aerators										
Maximum										
lbs	50.8	15.0	5.7	6.0	23.6	7.4	17.2	60.5	61.2	57.7
lbs/hr	0.770	0.315	0.120	0.125	0.262	0.414	0.716	1.008	1.274	1.201
lbs/1,000 gal	0.0507	0.0288	0.1291	0.0087	0.0137	0.0613	0.0653	0.0669	0.0712	0.0586
Minimum										
lbs	nil	nil	nil	nil	nil	nil	nil	nil	nil	nil
lbs/hr	nil	nil	nil	nil	nil	nil	nil	nil	nil	nil
Oxygen Transfer Rate Based on Retention Time										
ppm O_2/water	4.1	3.2	1.7	0.7	0.9	10.9	7.2	5.3	4.7	3.4
ppm/hr O_2/water	6.2	3.5	1.6	1.0	1.7	7.4	7.9	8.0	8.5	7.0
Oxygen Transferred as Required by Iron										
%	4.4	8.6	121.7	165.6	202.2	106.3	105.7	107.4	105.7	108.2
Oxygen Transferred as Aerator Rating										
%	2.6	1.1	0.4	0.4	0.9	1.4	2.4	3.4	4.2	4.0
Sludge Production										
gal/1,000 gal treated	12.5	15	20	20	35	--	145	95	112	120
solids content,mg/l	5200	4340	1400	2400	2770	--	2350	3460	2880	3390
lbs dry solids/1,000 gal treated	0.54	0.54	0.23	0.40	0.81	--	2.84	2.74	2.70	3.39

Table 4. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME - OXIDATION TANK AND SETTLING LAGOON ANALYSES (PROCESS NO. 10)

Test No.	9-70	10-70	3-70	2-70	1-70	8-70	4-70	5-70	6-70	7-70
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 1 & Proctor No. 2
pH, Av.	6.87	6.70	6.22	6.15	6.24	6.48	6.43	6.57	6.19	6.46
Range	6.60-7.36	6.60-7.10	5.80-6.68	4.50-6.43	5.57-6.61	6.44-6.50	5.98-6.72	6.14-6.87	5.29-6.75	6.25-6.55
Acidity, Av.	--	--	+36	+9	+14	--	+43	+23	+6	--
Range	--	--	+19 - +49	+5 - +13	-20 - +55	--	+41 - +45	-10 - +67	-15 - +25	--
Alkalinity, Av.	-6	-12	--	--	--	-29	--	--	--	-24
Range	-15 - +3	-15 - -6	--	--	--	-30 - -27	--	--	--	-40 - -12
Fe II, Av.	1	1	2	3	4	3	16	4	4	4
mg/l Range	--	--	1-3	2-4	1-7	--	3-28	2-6	2-7	2-6
Fe Total, Av.	--	--	--	--	6	--	--	--	--	--
mg/l Range	--	--	--	--	--	--	--	--	--	--
Suspended Solids, Av.	--	--	3	5	12	--	10	5	8	2
mg/l Range	--	--	2-3	4-6	11-12	--	--	3-7	--	1-3
Temperature, Av.	64	61	--	56	50	--	--	--	--	--
Range	62-68	60-63	--	53-60	47-53	--	--	--	--	--
Conductivity, Av.	759	948	580	519	822	1771	1488	1898	1958	1767
Range	668-836	900-1000	560-594	430-594	504-950	1680-1880	1370-1700	1736-1990	1900-2000	1580-1920
Effluent Rate, Av.	285	182	182	282	361	--	214	287	338	380
gpm Range	142-410	158-250	180-280	255-288	--	--	200-214	--	310-338	380-385
Turbidity, Av.	1	5	2	5	3	2	3	2	3	2
Range	1-7	4-7	2-3	2-22	2-10	2-3	2-4	2-3	2-4	1-12
Oxidation Tank Analysis										
pH, Av.	6.71	6.56	6.99	6.44	6.72	6.81	7.09	7.02	6.74	6.74
Range	6.29-7.77	6.13-7.17	6.50-7.98	6.25-6.50	5.73-8.53	6.71-7.04	6.94-7.29	6.16-7.88	5.80-7.54	6.58-7.03
Dissolved Oxygen, Av.	6.1	3.3	--	--	--	--	--	--	--	--
mg/l Range	4.5-7.8	2.1-4.5	--	--	--	--	--	--	--	--
Suspended Solids, ml/l										
Av.	65	--	28	48	97	--	341	324	339	406
Range	64-65	--	--	36-60	65-124	--	331-351	297-351	327-351	227-505

Table 5. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME - PROCESS CONDITIONS (PROCESS NO. 11)

Water Source	Average Flow, gpm	Run Duration hrs	Water Treated gal	Test No.
Bennett's Branch	103	45.5	281,440	71-71
Proctor No. 1	96	26.8	154,700	70-71
	100	35.0	210,000	38-71
Proctor No. 2	87	22.8	119,470	66-71
	89	19.5	104,980	56-71
	97	5.3	30,860	68-71
	100	19.8	119,140	67-71
	106	24.0	153,310	69-71
	134	7.0	56,370	57-71
	178	15.5	165,830	65-71

Process Flow:

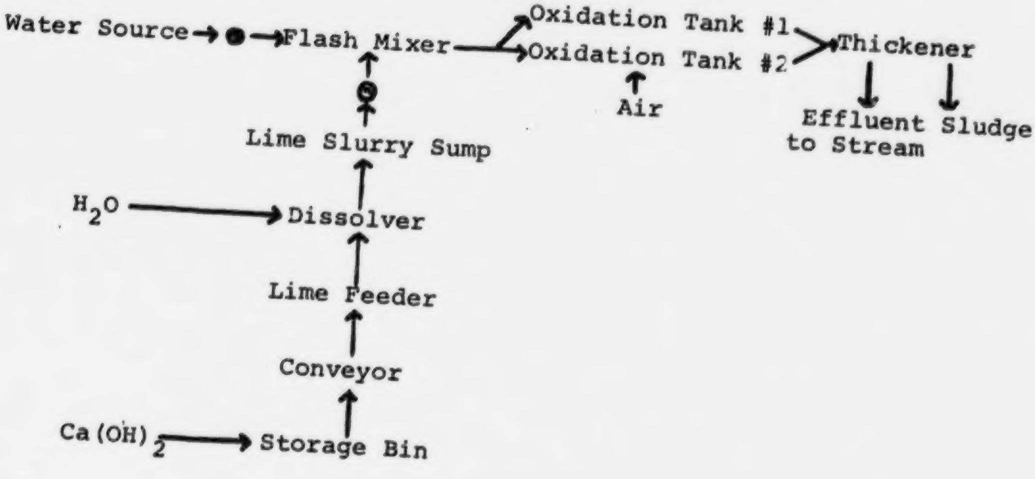

Water Source
Flash Mixer
Oxidation Tank #1
Oxidation Tank #2
Air
Thickener
Effluent to Stream
Sludge
Lime Slurry Sump
H_2O
Dissolver
Lime Feeder
Conveyor
$Ca(OH)_2$
Storage Bin

Table 7. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME - PROCESS AND REAGENT REQUIREMENTS

Test No.	71-71	38-71	70-71	56-71	65-71	66-71	67-71	68-71	69-71	57-71
Water Source	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2
Influent Fe II										
lbs	171.2	121.9	95.3	217.1	419.8	280.9	287.4	74.6	434.2	140.8
lbs/hr	3.763	3.483	3.555	11.136	27.081	12.322	14.514	14.078	18.090	20.119
Fe II Reacted										
lbs	168.9	106.3	93.9	217.1	418.4	280.0	286.4	74.4	431.6	140.4
Minimum Oxidation Rate										
mg/l/hr	71.9	60.7	72.8	247.8	303.2	280.8	288.1	289.5	337.5	298.7
Theoretical Reagent Required^1										
lbs	504^1	576^1	475^1	900^1	1689^2	1373^2	1246^2	328^2	1410^1	510^1
lbs/hr	11.1	12.7	17.7	46.1	109.0	60.2	62.9	60.9	58.8	72.9
lbs/1,000 gal	1.79	2.0	3.07	8.57	10.19	11.49	10.46	10.47	9.20	9.05
Reagent Utilized										
lbs	588	108.6	751	916	2852	3890	2714	727	2301	764
lbs/hr	12.9	31.0	28.0	47.0	184.0	170.6	137.1	137.2	95.9	109.1
lbs/1,000 gal	2.09	5.17	4.85	8.73	17.20	32.56	22.78	23.56	15.00	13.55
Alkali Consumption Utilized/ Theoretical										
%	116.6	18.9	158.1	101.8	168.9	283.3	217.8	221.6	163.2	149.8
Retention Time (minutes)										
Flash Mixer	14.4	14.8	15.4	16.6	8.3	17.0	14.8	15.3	14.0	11.1
Oxidation Tanks	192	199	206	223	112	228	199	204	187	148
Theoretical Oxygen Requirement										
lbs	24.5	17.5	13.6	31.1	60.1	40.2	41.2	10.7	62.2	20.2
lbs/hr	0.539	0.499	0.509	1.595	3.879	1.765	2.079	2.017	2.591	2.882
lbs/1,000 gal	0.0871	0.0833	0.0879	0.2962	0.3624	0.3365	0.3458	0.3467	0.4057	0.3583
ft^3 (std)	275.3	196.0	153.2	349.1	674.8	451.6	461.9	119.9	698.0	226.4
ft^3/hr (std)	6.05	5.60	5.72	17.90	43.54	19.81	23.33	22.63	29.08	32.34

Table 7 (continued). TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME - PROCESS AND REAGENT REQUIREMENTS

Test No.	71-71	38-71	70-71	56-71	65-71	66-71	67-71	68-71	69-71	57-71
Water Source	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2
Oxygen Utilized as Supplied by Aerators										
Maximum										
lbs	24.2	15.2	13.5	31.1	59.9	40.1	41.0	10.7	61.8	20.1
lbs/hr	0.533	0.435	0.502	1.595	3.866	1.759	2.072	2.010	2.576	2.872
lbs/1,000 gal	0.0860	0.0738	0.0873	0.2962	0.3612	0.3366	0.3441	0.3467	0.4031	0.3566
Minimum										
lbs	1.4	nil	0.9	22.7	47.6	31.3	32.0	8.4	49.4	15.5
lbs/hr	0.031	nil	0.035	1.162	3.072	1.371	1.617	1.588	2.058	2.215
Oxygen Transfer Rate Based on Retention Time										
ppm O_2/water	33.0	28.9	35.7	133.1	81.0	153.5	137.3	140.8	151.2	105.6
ppm/hr O_2/water	10.3	8.7	10.4	35.8	43.4	40.4	41.4	41.4	48.5	42.8
Oxygen Transferred as Required by Iron %	101.3	114.7	101.4	100.0	100.3	100.3	100.3	100.3	100.6	100.3
Oxygen Transferred as Aerator Rating %	0.9	0.7	0.8	2.7	6.4	2.9	3.5	3.4	4.3	4.8
Sludge Production										
gal/1,000 gal treated[3]	47	85	100	242	(260)	(260)	(260)	(260)	90 (10 hrs)	(260)
solids content, mg/l	4250	6900	4960	5780	5150	6540	6580	5760	26500	5760
lbs dry solids/1,000 gal treated	1.67	4.90	4.14	11.70	11.20	14.15	14.25	12.50	15.50	12.50

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1. Based on 93.98% $Ca(OH)_2$
2. Bulk hydrated lime had become carbonated during storage. Assume 81.43% $Ca(OH)_2$ equiv.
3. Two hour settling time.

Table 8. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED LIME - OXIDATION TANK ANALYSES (PROCESS NO. 11)

Test No.	71-71	38-71	70-71	56-71	65-71	66-71	67-71	68-71	69-71	57-71[1]
Water Source	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2
pH, Av.	8.58	8.47	8.32	7.94	7.22	9.11	8.80	8.25	8.18	7.05
Range	7.15-9.43	7.73-9.30	6.78-9.30	7.28-8.38	6.00-8.05	7.05-10.05	8.54-9.09	8.20-8.30	7.03-8.65	
Acidity or Alkalinity*										
mg/l, Av.	+18	-33	+26	-17	+8	+21	-11	-12	-6	-3
Range	-23 - +49	-345 - +60	+15 - +49	-7 - -24	--	+1 - +74	-19 - -2	--	-40 - +15	--
Iron II, Av.	1	4	1	1	1	1	1	1	2	1
mg/l Range	--	1-9	--	--	--	1-2	--	--	1-4	--
Oxidation-Reduction Potential										
milliv, Av.	162	350	166	190	275	176	169	200	--	178
Range	110-215	302-400	130-215	162-200	250-295	130-250	138-200	--	--	136-210
Suspended Solids, Av.	200	588	496	--	1340	1700	1710	--	1860	--
ml/l Range	188-212	328-848	--	--	--	--	--	--	--	--
			Thickener Effluent Analyses							
pH, Av.	--	7.57	--	7.57	9.69	9.77	9.75	--	--	7.60
Range	--	7.55-7.62	--	7.54-7.67	9.65-9.75	--	9.70-9.79	--	--	7.53-7.74

*Negative value is Alkalinity.

[1] Average values (mg/l)
Sulfate - 1542
Calcium - 568
Magnesium - 48

Table 9. TREATMENT OF COAL MINE DRAINAGE WITH CALCINED LIME - PROCESS CONDITIONS (PROCESS NO. 10)

Water Source	Average Flow gpm	Run Duration hrs	Water Treated gal	Test No.
Tyler Run	101	23.5	143,220	16-71
	199	37.0	443,770	14-71
Bennett's Branch	109	45.3	297,500	8-71
	197	62.5	741,490	7-71
	289	60.0	1,040,560	6-71
Proctor No. 1	107	21.5	138,030	9-71
	196	45.3	534,960	10-71
Proctor No. 2	103	40.3	250,590	11-71
	195	29.2	342,900	13-71
	199	41.0	490,210	12-71
	200	50.0	601,330	15-71
Proctor No. 1 and Proctor No. 2	325 (260+65)	29.3	571,560	17-71

Process Flow:

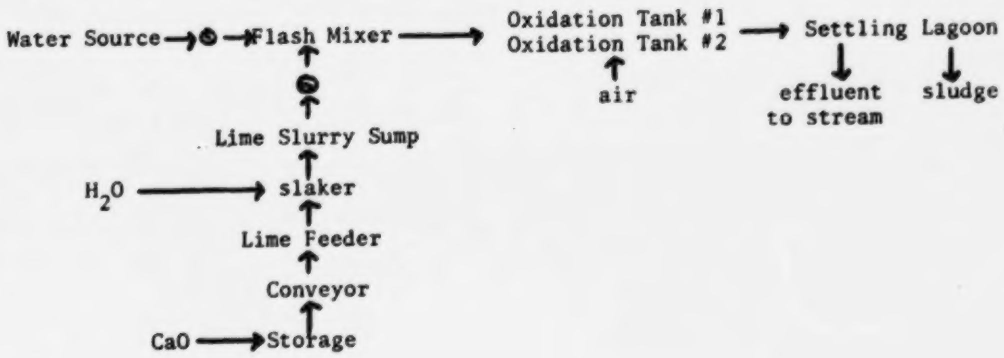
Water Source
Flash Mixer
Oxidation Tank #1
Oxidation Tank #2
air
Settling Lagoon
effluent to stream
sludge
Lime Slurry Sump
H_2O
slaker
Lime Feeder
Conveyor
CaO
Storage

Table 10. TREATMENT OF COAL MINE DRAINAGE WITH CALCINED LIME - RAW WATER ANALYSES

Test No.	16-71	14-71	8-71	7-71	6-71	9-71	10-71	11-71	13-71	12-71	15-71	17-71
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor *1 & Proctor *2
Temp. °F., Av.	37	41	34	34		46	45	52			45	36
Range	35-42	40-43				45-47	43-49	51-52			40-54	35-37
pH, Av.	3.82	4.02	3.94	3.04	3.02	3.87	3.47	3.02	3.52	3.95	2.92	2.81
Range	3.79-3.89	3.78-4.23	3.51-4.37	2.40-3.92	2.99-3.05	2.99-5.13	3.36-3.80	2.38-3.79	3.51-3.54	3.11-4.82	2.72-3.05	2.68-3.07
Acidity mg/l, Av.	74	73	161	140	281	477	501	1736	1603	1580	1390	1518
Range	-	70-79	114-208	106-176	202-430	475-479	493-507	1652-1848	1443-1762	1478-1696	-	-
Iron II,mg/l, Av.	1	1	32	26	32	83	84	419	82	263	239	320
Range	-	-	29-35	25-27	26-35	81-85	80-86	413-421	80-83	147-402	212-261	-
Iron III, mg/l Av.	3*	3*	≤0*	5*	2*	≤0*	≤0*	25*	350*	174*	173*	5*
Al, mg/l, Av.	13*	14*	21*	22*	23*	79*	78*	192*	196*	195*	200*	179*
Ca, mg/l, Av.	59*	59*	36*	36*	36*	96*	96*	201*	201*	201*	201*	125*
Mg, mg/l, Av.	64*	64*	40*	40*	40*	76*	76*	97*	97*	97*	97*	85*
SO_4, mg/l, Av.	171*	219*	293*	310*	332*	784*	782*	2404*	2356*	2369*	2318*	1961*
Conductivity, micromhos												
Av.	645	660	1152	559*	592*	1238*	1236*	1800	2689*	2704*	1833	680
Range	640-660	640-680	1100-1220								1800-1900	-

*Values established from monthly stream analyses.

Table 11. TREATMENT OF COAL MINE DRAINAGE WITH CALCINED LIME - PROCESS AND REAGENT REQUIREMENTS

Test No.	16-71	14-71	8-71	7-71	6-71	9-71	10-71	11-71	13-71	12-71	15-71	17-71
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor 1 & 2
Influent Fe II												
lbs.	1.2	3.7	79.4	153.8	277.7	95.3	374.9	874.8	261.9	1073.0	1196.2	1522.4
lbs./hr.	0.51	0.100	1.746	2.563	4.628	4.445	8.240	21.599	8.955	26.195	27.925	52.050
Fe II Reacted												
lbs.	1.2	3.7	71.4	142.2	234.6	95.7	370.4	870.4	229.9	1068.1	1196.2	1517.7
Minimum Oxidation Rate, mg/l/hr.	1.0	1.0	28.8	23.0	27.0	83.2	83.0	416.5	80.4	261.1	238.3	318.2
Theoretical Reagent Required[1]												
lbs.	51.7	157.8	233.3	505.6	1431.0	320.7	1305.4	2118.9	2677.2	3772.5	4071.1	4225.9
lbs./hr.	2.20	4.26	5.15	8.09	23.85	14.9	28.82	52.58	91.68	92.01	81.42	144.2
lbs./1,000 gal.	0.36	0.36	0.78	0.68	1.38	2.32	2.14	8.46	7.81	7.70	6.77	7.39
Reagent Utilized												
lbs.	40	227	380	750	1220	500	2160	3700	6707	10298	8827	6869
lbs./hr.	1.70	6.14	8.39	12.00	20.33	23.26	47.68	91.81	229.69	251.17	176.54	234.13
lbs./1000 gal.	0.28	0.51	1.28	1.01	1.17	3.62	4.04	14.77	19.56	21.01	14.68	12.00
Alkali Consumption Utilized / Theoretical %	77.5	143.8	163.1	148.2	85.3	156.1	165.3	175.0	251.5	272.2	217.0	161.7
Alkali Consumption Theor. + Effl. Alk. / Utilized %			55	52	113		57	57	41	37		
Retention Time (min.)												
Flash Mixer	14.7	7.4	13.6	7.5	5.1	13.4	7.6	14.4	7.6	7.4	7.4	4.6
Oxidation Tanks	196	100	182	101	68.6	186	101	193	102	100	99	61.1
Theoretical Oxygen Requirement												
lbs.	0.2	0.5	11.4	22.0	41.6	13.7	53.7	125.3	37.5	153.5	171.4	218.1
lbs./hr.	0.007	0.014	0.250	0.367	0.693	0.637	1.180	3.094	1.283	3.743	3.427	7.456
lbs./1,000 gal.	0.0014	0.0011	0.0383	0.2966	0.0400	0.0992	0.1004	0.5000	0.1094	0.3131	0.2850	0.3816
ft.^{3} (std.)	1.9	5.9	127.7	247.3	466.7	153.7	602.7	1406.3	421.1	1722.3	1922.9	2448.0
ft.^{3}/hr. (std.)	0.08	0.16	2.81	4.12	7.78	7.15	13.246	34.724	14.40	42.01	38.45	83.68
Oxygen Utilized as Supplied by Aerators												
Maximum												
lbs.	13.4	40.7	12.0	20.4	17.1	13.7	84.0	124.6	58.9	195.1	213.7	252.2
lbs./hr.	0.568	1.100	0.263	0.340	0.474	0.744	1.846	3.077	2.013	1.758	4.273	8.620
lbs./1,000 gal.	0.0094	0.0917	0.0403	0.2751	0.0164	0.0992	0.1570	0.4972	0.1718	0.3980	0.3554	0.4412
Minimum												
lbs.	nil	17.2	nil	nil	nil	1.4	35.4	103.76	30.1	153.0	162.8	206.0
lbs./hr.	nil	0.465	nil	nil	nil	0.067	0.777	2.562	1.027	3.732	3.26	7.012
Oxygen Transfer Rate Based on Retention Time												
ppm O_2/water	36.6	18.3	14.6	5.7	3.8	43.1	31.6	192.0	35.0	79.7	30.4	54.0
ppm/hr. O_2/water	11.2	11.0	4.8	3.4	3.3	13.9	18.8	59.7	20.6	47.8	42.7	53.0
Oxygen Transferred as Required by Iron, %	1.2	1.3	95.1	107.9	146.2	83.6	63.9	100.6	63.7	78.7	80.2	86.5
Oxygen Transferred as Aerator Rating, %	0.9	1.8	0.4	0.6	0.8	1.2	3.1	5.1	3.4	7.9	7.1	14.4
Sludge Production												
gal./1,000 gal. treated[2]	(13)	(13)	35	25	23	--	200	296	(242)	(250)	(250)	(250)
solids content, mg/l	5000	5000	4720	3240	4260	--	2000	6450	9450	5920	6000	6000
lbs. dry solids/ 1,000 gal. treated	0.55	0.55	1.38	0.68	0.82	--	3.58	15.90	19.10	12.35	12.50	12.50

[1] Based on 96.0% CaO

[2] Two hour settling time

Table 12. TREATMENT OF COAL MINE DRAINAGE WITH CALCINED LIME - OXIDATION TANK ANALYSES (PROCESS NO. 10)

Test No.	16-71	14-71	8-71	7-71	6-71	9-71	10-71	11-71	13-71	12-71	15-71	17-71
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 1 & 2
pH, Av.	9.05	9.34	6.85	7.03	7.53	6.95	7.81	8.14	8.96	8.79	8.50	7.70
Range	6.74-11.69	8.31-11.01	5.87-8.25	6.05-7.74	6.41-8.88	6.01-8.04	5.99-9.40	7.45-8.93	7.90-9.52	7.22-9.52	4.44-10.39	4.21-9.48
Acidity or Alkalinity,mg/1*												
Av.	+14	-7	--	--	-55	--	+3	+2	--	-16	-6	-13
Range	-6 - +20	-15 - -1	--	--	-176 - +22	--	-15 - +28	-63 - +178	--	-72 - +18	--	-36 - +11
Fe II, Av.	--	1	--	1	5	--	2	11	1	1	--	1
mg/1 Range	--	--	--	--	1-13	--	0-10	0-29	--	0-1	--	--
Suspended Solids mg/1, Av.	--	--	165	81	98	--	430	1903	2289	1480	--	--
Range	--	--	--	--	96-99	--	--	--	--	--	--	--
Dissolved Oxygen mg/1, Av.	11.1	10.9	0.4	--	0.4	0.3	7.3	--	9.1	10.3	8.4	7.3
Range	8.0-12.0	9.2-11.4	--	--	--	0.2-0.3	0.2-15.0	--	--	--	7.4-11.2	4.7-11.9
Oxidation-Reduction Potential,milliv												
Av.	215	198	361	--	371	414	327	295	213	208	254	271
Range	170-260	110-285	260-405	--	280-440	345-443	200-475	240-330	212-214	180-214	175-525	170-580

* Negative value is alkalinity.

Table 13. TREATMENT OF COAL MINE DRAINAGE WITH CALCINED LIME - SETTLING LAGOON EFFLUENT ANALYSES (PROCESS NO. 10)

Test No.	16-71	14-71	8-71	7-71	6-71	9-71	10-71	11-71	13-71	12-71	15-71	17-71
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 2	Proctor No. 1 & 2
pH, Av.	8.36	9.44	7.54	7.34	7.32	8.06	8.01	8.23	9.40	9.14	—	7.18
Range	6.50-9.67	8.90-10.61	7.00-8.93	6.67-9.67	6.18-11.38	5.95-9.37	6.75-9.51	7.45-9.51	9.20-9.62	8.75-9.53	—	4.23-9.42
Acidity, Av.	—	—	+17	+32	+13	—	+29	+10	-33	-9	—	—
mg/1 Range	—	—	-24 - +64	-1 - +64	-1 - +26	—	+3 - +64	+2 - +20	-15 - -50	-33 - +25	—	—
Fe II, Av.	—	—	1	3	4	13	1	0	1	1	—	—
mg/1 Range	—	—	—	1-4	2-6	—	—	—	—	0-2	—	—
Fe Total, mg/1, Av.	—	—	1	—	5	—	1	1	—	1	—	—
Suspended Solids mg/1, Av.	—	—	6	10	9	—	6	6	3	23	—	—
Range	—	—	—	—	—	—	5-7	1-13	—	9-44	—	—
Temp. °F, Av.	35	41	32	32	32	35	38	40	43	39	41	41
Range	34-41	38-47	—	—	—	32-37	37-39	38-41	41-47	36-41	40-44	34-45
Conductivity, Av.	1180	686	1115	1014	1290	1684	—	—	—	—	—	—
Range	840-1760	560-800	1100-1140	940-1120	1160-1760	1140-2020	—	—	—	—	—	—
Effluent Rate gpm Av.	105	256	127	—	380	106	273	125	249	297	266	370
Range	105-206	—	83-218	—	375-405	100-110	210-273	67-163	248-250	243-297	203-370	145-445
SO_4, mg/1, Av.	—	—	275	—	296	—	698	—	—	—	—	—
Calcium, mg/1, Av.	—	—	—	94	94	—	246	—	—	—	—	—

Table 14. Treatment of Coal Mine Drainage with Sodium Hydroxide Process Conditions (Process No. 10)

Water source	Average flow, gpm	Run duration hrs	Water treated gal	Test no.
Tyler Run	91	48.0	263,230	1-71
Bennett's Branch	87	30.0	157,280	2-71
	199	48.0	573,900	3-71
	292	24.0	421,380	4-71
Proctor No. 1	88	36.0	190,570	5-71
	204	21.5	263,160	18-71
	255	24.0	368,250	19-71
Proctor No. 2	87	13.5	70,880	20-71
	195	25.3	296,110	21-71
	227	27.0	367,760	22-71

Water Flow:

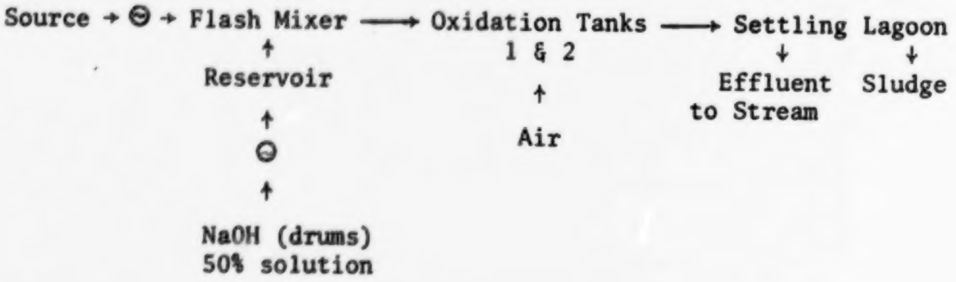
Source
Flash Mixer
Oxidation Tanks
1 & 2
Settling Lagoon
Reservoir
Air
Effluent
to Stream
Sludge
NaOH (drums)
50% solution

Table 15. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM HYDROXIDE - RAW WATER ANALYSES (PROCESS NO. 10)

Test No.	1-71	2-71	3-71	4-71	5-71	18-71	19-71	20-71	21-71	22-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2
Temp., Av.	34	35	34	34	45	46	46	41	51	46
°F Range	--	34-35	--	--	44-46	45-47	44-48	--	--	--
pH, Av.	3.97	4.19	4.49	4.38	3.50	3.50	3.49	3.00	2.84	2.83
Range	3.47-4.43	4.02-4.47	4.40-4.55	4.15-4.58	3.12-3.79	3.47-3.53	3.45-3.51	2.94-3.08	2.83-2.87	2.80-2.66
Acidity, Av.	123	138	129	98	591	376	403	1142	1616	1533*
mg/l Range	114-140	117-152	62-172	87-109	516-660	302-450	386-419	1250-1634	--	--
Fe II, Av.	1*	28*	32	24	78	73	39	331	307	316*
mg/l Range	--	26-30	29-36	20-29	72-84	72-74	--	--	--	--
Fe III, mg/l, Av.	3*	≥0*	≥0*	6*	≥0*	≥0*	32*	56*	53*	40*
Al, mg/l, Av.	16*	18*	19*	20*	80*	74*	74*	207*	212*	218*
Ca, mg/l, Av.	59*	36*	36*	36*	96*	96*	96*	201*	201*	201*
Mg, mg/l, Av.	64*	40*	40*	40*	76*	76*	76*	97*	97*	97*
SO_4, Av.	962	966	1091	1012	1885	753*	753*	2800	2577	2540*
mg/l Range	880-1000	960-1000	1040-1160	940-1120	1600-2400	--	--	1800-3200	2200-3700	--
Conductivity, Av. micromhos	374*	592*	592*	592*	1260*	1195*	1195*	2580*	2580*	2580*

*Values established from monthly analyses.

Table 16. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM HYDROXIDE - OXIDATION TANK ANALYSES (PROCESS NO. 10)

Test No.	1-71	2-71	3-71	4-71	5-71	18-71	19-71	20-71	21-71	22-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2
pH, Av.	6.29	6.12	6.67	6.93	6.74	9.03	8.01	9.05	7.42	6.94
Range	5.15-8.70	5.63-6.72	5.60-9.00	6.55-7.32	6.11-7.40	7.33-10.00	7.52-9.10	8.25-10.00	4.33-8.60	5.18-8.05
Acidity or Alkalinity * mg/l, Av.	+18	--	+29	--	--	-43	+9	-35	-72	-49
Range	--	--	--	--	--	-58 - -21	--	-11 - -58	-118 - -26	-75 - -23
Iron II, Av.	--	--	--	--	5	1	1	1	11	1
mg/l Range	--	--	--	--	1-11	--	--	--	1-27	--
Suspended Solids, Av.	44	86	91	106	259	--	393	1530	1210	--
ml/l Range	39-48	--	85-95	--	--	--	--	--	--	--
Dissolved Oxygen ppm, Av.	--	0.4	0.4	0.4	0.2	10.0	10.3	8.8	5.5	5.6
Range		0.3-0.4	0.3-0.4	0.3-0.4	0.2-0.3	9.8-10.2	10.1-10.5	8.2-10.4	4.8-5.8	4.9-8.3
Oxidation-Reduction Potential, milliv. Av.	540	416	432	414	536	--	262	218	128	329
Range	525-581	380-450	400-480	400-420	527-544	--	214-309	180-272	40-180	214-560

*Negative value is Alkalinity.

Table 17. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM HYDROXIDE - PROCESS AND REAGENT REQUIREMENTS

Test No.	1-71	2-71	3-71	4-71	5-71	18-71	19-71	20-71	21-71	22-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2
Influent Fe II										
lbs.	2.2	36.9	167.8	82.4	123.7	160.2	220.5	188.1	813.2	1012.2
lbs./hr.	0.046	1.219	3.187	3.508	3.435	7.453	9.189	13.933	32.205	37.490
Fe II Reacted										
lbs.	2.2	36.9	123.9	61.8	117.7	160.2	220.5	187.5	776.2	1009.2
Minimum Oxidation Rate										
mg/l/hr.	1.00	28.1	25.9	17.6	74.0	73.0	71.8	316.7	314.3	328.8
Theoretical Reagent Required										
lbs.	216.0	144.8	493.9	275.5	751.4	660.2	990.1	540.1	3192.6	3761.5
lbs./hr.	4.50	4.83	10.29	11.48	20.87	30.71	41.25	40.0	126.19	139.31
lbs./1,000 gal.	0.82	0.92	0.86	0.65	3.94	2.51	2.69	7.62	10.78	10.23
Reagent Utilized										
lbs.	178.6	95.7	350.9	344.5	618.9	1154.8	1295.1	880.4	4561.7	5946.2
lbs./hr.	3.72	3.19	3.71	14.35	17.19	53.71	53.96	65.22	180.31	220.23
lbs./1,000 gal.	0.68	0.61	0.61	0.83	3.25	4.39	3.52	12.42	15.41	16.17
Alkali Consumption Utilized/Theoretical										
%	82.7	66.1	71.0	125.0	82.4	174.9	130.8	163.0	142.9	158.1
$\frac{\text{Theor.} + \text{Effl. Alk.}}{\text{Utilized}}$										
%	96	241	97	52	118		74	64	73	
Retention Time (minutes)										
Flash Mixer	16.3	17.0	7.4	5.1	16.8	7.3	5.8	17.0	7.6	6.5
Oxidation Tanks	218	228	100	68	226	97	78	228	102	87

Table 17 (continued). TREATMENT OF COAL MINE DRAINAGE WITH SODIUM HYDROXIDE - PROCESS AND REAGENT REQUIREMENTS

Test No.	1-71	2-71	3-71	4-71	5-71	18-71	19-71	20-71	21-71	22-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2
Theoretical Oxygen Requirement										
lbs.	0.3	5.3	24.0	11.8	17.7	23.0	31.6	26.9	116.5	144.9
lbs./hr.	0.007	0.175	0.457	0.502	0.492	1.068	1.316	1.996	4.613	5.370
lbs./1,000 gal.	0.0011	0.0337	0.0418	0.0280	0.0929	0.0874	0.0858	0.3795	0.3934	0.3940
ft.^{3} (std.)	3.8	59.3	269.7	132.4	198.8	257.6	354.5	302.4	1307.3	1627.2
ft.^{3}/hr. (std.)	0.08	1.96	5.13	5.63	5.52	11.98	14.77	22.40	51.77	60.27
Oxygen Utilized as Supplied by Aerators										
Maximum										
lbs.	0.3	5.5	19.5	10.0	17.1	46.0	63.1	32.0	124.7	161.7
lbs./hr.	0.007	0.183	0.408	0.426	0.474	2.140	2.627	2.373	4.940	5.990
lbs./1,000 gal.	0.0011	0.0350	0.340	0.0237	0.0897	0.1748	0.1714	0.4515	0.4211	0.4397
Minimum										
lbs.	nil	nil	nil	nil	nil	44.1	30.3	25.3	100.4	131.1
lbs./hr.	nil	nil	nil	nil	nil	2.053	1.264	1.875	3.975	4.855
Oxygen Transfer Rate Based on Retention Time										
ppm O_2/water	0.5	16.0	6.7	3.3	40.7	8.6	26.8	207.1	86.0	76.4
ppm/hr. O_2/water	0.15	4.2	4.0	2.9	10.8	5.3	20.6	54.5	50.6	52.7
Oxygen Transferred as Required by Iron										
%	100.0	95.6	112.0	117.8	103.8	49.9	50.1	84.1	93.4	89.6
Oxygen Transferred as Aerator Rating										
%	0.01	0.3	0.7	0.7	0.8	3.6	4.4	4.0	8.2	10.0
Sludge Production										
gal./1,000 gal.[a] treated	30	35	(35)	(35)	(140)	--	(140)	(200)	(200)	--
solids content, mg/l	1465	2460	2600	3020	1850	--	2800	7650	6050	--
lbs. dry solids/ 1,000 gal. treated	0.367	0.717	0.760	0.882	2.16	--	3.28	12.75	10.10	--

[a] Two hour settling time.

Table 18. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM HYDROXIDE - SETTLING LAGOON EFFLUENT ANALYSES (PROCESS NO. 10)

Test No.	1-71	2-71	3-71	4-71	5-71	18-71	19-71	20-71	21-71	22-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2	Proctor No. 2
pH, Av.	6.76	8.75	6.75	6.84	6.74	8.37	8.54	8.92	6.31	5.48
Range	6.32-7.60	6.98-10.46	6.28-7.15	6.59-7.00	6.40-6.92	6.88-9.18	8.00-9.23	8.47-9.30	5.75-7.59	5.14-5.88
Acidity, Av.	+26	--	+41	+33	+16	--	+15	--	--	--
mg/l Range	-8 - +45	--	+28 - +62	+11 - +64	--	--	+7 - +22	--	--	--
Alkalinity, Av.	--	-83	--	--	--	--	--	-49	-70	--
mg/l Range	--	--	--	--	--	--	--	-92 - -6	--	--
Iron II, Av.	0	0	6	5	3	--	0	1	0	--
mg/l Range	--	--	2-11	4-6	2-6	--	--	0-1	--	--
Iron Total, Av. mg/l	1	1	--	7	--	--	1	1	0	--
Suspended Solids, Av.	4	10	5	--	--	7	22	5	29	--
mg/l Range	1-5	9-11	--	--	--	--	3-41	--	--	--
Temp., Av.	32	32	32	32	33	--	--	--	--	--
°F Range	--	--	--	--	32-35	--	--	--	--	--
Conductivity, Av.	966	1013	1046	1016	1428	--	--	--	--	--
micromhos Range	940-1040	980-1040	1000-1080	1000-1040	1000-1920	--	--	--	--	--
Effluent Rate, Av.	70	123	232	351	92	302	329	108	128	60
gpm Range	42-107	95-123	185-270	351-380	40-100	285-302	300-380	97-320	40-180	58-61
Tubidity, Av.	2	--	--	--	--	10	3	8	--	--
jtu Range	2-3	--	--	--	--	7-13	2-6	2-15	--	--
SO_4, mg/l, Av.	212	234								

Table 19. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE - PROCESS CONDITIONS (PROCESS NO. 10)

Water Source	Average Flow, gpm	Run Duration hrs	Water Treated gal	Test No.
Tyler Run	276	24.0	398,150	25-71
Bennett's Branch	100	22.8	137,869	30-71
	144	14.3	123,800	27-71
	196	25.0	294,880	29-71
	288	24.0	415,000	26-71
Proctor No. 1	106	23.8	152,370	32-71
	169	24.3	246,731	31-71
	222	45.5	607,300	28-71
Proctor No. 2	105	21.8	137,860	33-71
	187	12.8	144,200	34-71

Water Flow:

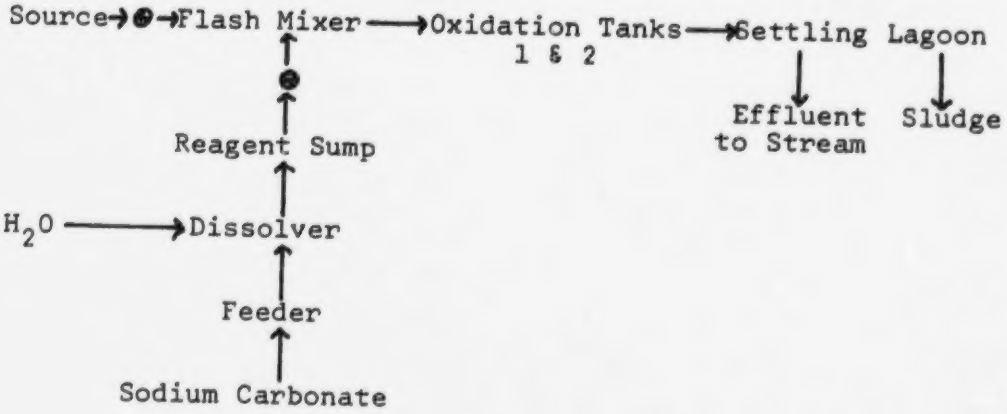
Source
Flash Mixer
Oxidation Tanks
1 & 2
Settling Lagoon
Effluent
to Stream
Sludge
Reagent Sump
H_2O
Dissolver
Feeder
Sodium Carbonate

Table 20. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE - RAW WATER ANALYSES (PROCESS NO. 10)

Test No.	25-71	26-71	29-71	27-71	30-71	28-71	31-71	32-71	34-71	33-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2
Temp., Av.	38	36	41	41	41	46	45	45	50	50
°F Range	26-41	34-38	--	--	--	43-50	--	--	--	--
pH, Av.	3.76	4.09	4.86	4.26	4.73	3.68	3.55	3.50	2.83	2.92
Range	3.70-3.80	3.79-4.21	4.48-5.72	4.02-4.62	4.69-4.80	3.50-4.03	3.51-3.60	3.48-3.53	2.78-2.87	2.91-2.93
Acidity, Av.	71	84	62	130	72	417	332	567	1243	1398
mg/l Range	69-72	82-86	52-69	86-173	51-86	397-440	328-348	230-1434	1174-1311	--
Fe II, Av.	1	8	14	28	15	61	57	52	118	241
mg/l Range	--	7-10	14-14.5	20-35	12-18	50-67	56-58	45-59	--	--
Fe III, mg/l, Av.	3*	9*	3*	>0*	2*	10*	20*	20*	155*	126
Al, mg/l, Av.	14*	18*	18*	18*	19*	65*	63*	62*	245*	240*
Ca, mg/l, Av.	59*	36*	36*	36*	36*	96*	96*	96*	201*	201*
Mg, mg/l, Av.	64*	40*	40*	40*	40*	76*	76*	76*	97*	97*
SO_4, mg/l, Av.	131	141	130*	150*	114	520*	550	1163	1628	1942*
Range	--	--	--	--	--	--	--	486-1840	--	--
Conductivity, Av.	320*	604	532	600	484	300*	1926	1906	270*	2312
micromhos Range	--	560-680	500-580	560-640	480-500	--	1920-1940	1840-1980	--	1800-2550

*Values established from monthly analyses.

Table 21. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE - PROCESS AND REAGENT REQUIREMENTS

Test No.	25-71	26-71	29-71	27-71	30-71	28-71	31-71	32-71	34-71	33-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2
Influent Fe II										
lbs.	3.3	34.6	35.1	29.3	17.3	308.4	115.7	65.5	141.7	275.5
lbs./hr.	0.138	1.441	1.403	2.018	0.716	6.777	4.821	2.759	11.043	12.664
Fe II Reacted										
lbs.	3.3	31.1	28.5	28.2	10.9	303.3	113.7	64.3	139.1	264.6
Minimum Oxidation Rate										
mg/l/hr.	1.0	9.0	11.6	27.4	9.5	59.8	55.2	50.6	115.6	230.0
Theoretical Reagent Required										
lbs.	249.8	308.1	161.6	142.2	87.6	2238.1	723.9	763.5	1584.1	1703.3
lbs./hr.	10.1	12.85	6.47	9.94	3.85	49.19	29.79	32.08	123.76	78.13
lbs./1,000 gal.	0.63	0.74	0.55	1.15	0.64	3.69	2.93	5.01	10.99	12.36
Reagent Utilized										
lbs.	562	963	375	304	145	6004	2011	1117	4648	4542
lbs./hr.	23.42	40.13	15.0	21.26	6.36	131.96	82.76	46.93	363.13	208.35
lbs./1,000 gal.	1.41	2.32	1.27	2.46	1.05	9.89	8.15	7.33	32.23	32.95
Alkali Consumption Utilized/Theoretical										
%	225.0	312.6	232.1	213.8	165.5	268.3	277.8	146.3	293.4	266.7
$\frac{\text{Alkali Consumption Theor. + Effl. Alk.}}{\text{Utilized}}$										
%	108	49	79	70	75	50	47	83	41	41
Retention Time (minutes)										
Flash Mixer	5.4	5.1	7.6	10.3	14.8	6.7	8.8	14.0	7.9	14.1
Oxidation Tanks	72	69	103	138	199	89	118	187	106	189

Table 21 (continued). TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE - PROCESS AND REAGENT REQUIREMENTS

Test No.	25-71	26-71	29-71	27-71	30-71	28-71	31-71	32-71	34-71	33-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2
Theoretical Oxygen Requirements										
lbs.	0.5	5.0	5.0	4.2	2.5	44.2	16.6	9.4	20.3	39.5
lbs./hr.	0.020	0.206	0.201	0.289	0.109	0.971	0.691	0.395	1.582	1.814
lbs./1,000 gal.	0.0012	0.0120	0.0169	0.0339	0.0181	0.0728	0.0673	0.0617	0.1408	0.2865
$ft.^3$ (std.)	5.3	55.6	56.4	47.0	27.8	495.7	186.0	105.3	227.8	442.8
$ft.^3$/hr. (std.)	0.222	2.35	2.26	3.24	1.22	10.89	7.75	4.43	17.75	20.35
Oxygen Utilized as Supplied by Aerators										
Maximum										
lbs.	20.7	25.2	14.6	10.3	6.5	65.7	21.7	11.6	33.6	39.4
lbs./hr.	0.861	1.051	0.585	0.711	0.284	1.444	0.907	0.488	2.619	1.811
lbs./1,000 gal.	0.0520	0.0607	0.0495	0.0832	0.0471	0.1082	0.0879	0.0761	0.2330	0.2858
Minimum										
lbs.	nil	nil	nil	nil	nil	11.6	nil	nil	11.1	27.8
lbs./hr.	nil	nil	nil	nil	nil	0.255	nil	nil	0.861	1.276
Oxygen Transfer Rate Based on Retention Time										
ppm O_2/water	7.4	8.4	10.3	22.8	18.9	19.3	21.0	28.7	49.5	108.4
ppm/hr. O_2/water	6.2	7.3	6.0	9.9	5.7	13.0	10.7	9.2	28.0	34.4
Oxygen Transferred as Required by Iron %	2.3	19.6	34.4	40.6	38.4	67.2	62.7	80.9	60.4	100.1
Oxygen Transferred as Aerator Rating %	1.4	1.8	1.0	1.2	0.5	2.4	1.5	0.8	4.4	3.0
Sludge Production										
gal./1,000 gal. treated[1]	3	10	(10)	(10)	(10)	50	30	30	80	110
solids content, mg/l	10600	50700	4400	6000	3700	4400	7000	5940	10000	21100
lbs. dry solids/ 1,000 gal. treated	0.250	0.476	0.367	0.500	0.309	1.84	1.75	1.48	6.67	19.35

[1] Two hour settling time

Table 22. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE - OXIDATION TANK ANALYSES (PROCESS NO. 10)

Test No.	25-71	26-71	29-71	27-71	30-71	28-71	31-71	32-71	34-71	33-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2
pH, Av.	7.96	7.74	7.70	7.95	7.44	7.55	7.49	7.80	7.44	7.31
Range	7.24-10.00	7.40-8.08	6.82-7.03	7.56-8.21	7.12-8.00	7.24-8.01	7.00-7.85	7.56-8.08	7.31-7.73	6.90-7.74
Acidity or Alkalinity[*] mg/l, Av.	-81	-43	-2	-50	0	-135	-137	--	-39	-432
Range	-212- -40	-33 - -58	-12 - +8	-33 - -66	-26 - +26	-120 - -147	-136 - -138	--	--	-293 - -570
Fe II, Av.	--	6	3	1	6	1	1	--	--	9
mg/l Range	--	--	-12 - +8	--	--	--	--	--	--	1-18
Fe III, mg/l, Av.	1	1	--	--	--	1	0	--	1	--
Suspended Solids, Av.	31	57	44	60	37	220	210	178	800	2318
mg/l Range	--	--	--	--	--	--	--	156-200	--	--
Dissolved Oxygen, Av.	6.1	6.0	4.4	6.0	4.3	4.3	2.7	1.9	2.8	1.2
ppm Range	5.9-6.4	4.8-6.2	1.4-5.2	5.8-6.3	4.0-5.0	3.4-6.0	2.1-4.0	1.8-2.1	2.6-2.9	1.0-1.8
Oxidation-Reduction Potential, milliv. Av.	332	284	383	313	357	198	85	154	100	220
Range	240-370	260-360	370-420	270-340	340-370	75-300	80-90	100-215	--	--

[*] Negative value is Alkalinity.

Table 23. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE - SETTLING LAGOON EFFLUENT ANALYSES (PROCESS NO. 10)

Test No.	25-71	26-71	29-71	27-71	30-71	28-71	31-71	32-71	34-71	33-71
Water Source	Tyler Run	Bennett's Branch	Bennett's Branch	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1	Proctor No. 1	Proctor No. 2	Proctor No. 2
pH, Av.	8.58	7.79	7.26	7.84	7.52	7.71	7.53	7.51	7.55	7.29
Range	7.62-9.78	7.5-8.14	6.94-7.78	7.68-8.15	6.95-8.10	7.45-8.10	7.28-8.02	7.25-7.84	7.17-8.29	6.95 - 8.11
Alkalinity, Av.	-101	-45	-51	-64	-17	-137	-105	-124	-262	-148
Range	-34 - -195	-58 - -24	-100 - -1	-92 - -35	-35 - +1	-155 - -120	-120 - -76	-138 - -110	-397 - -127	-170 - -120
Fe II, mg/l, Av.	--	1	1	--	--	1	1	1	1	2
Fe Total, mg/l, Av.	≧1	2	--	--	≧1	2	2	2	1	3
Suspended Solids, Av.	3	5	3	6	7	7	7	16	26	4
mg/l Range	--	2-7	--	--	5-12	4-10	3-10	3-38	--	--
Temp., Av.	37	35	39	34	37	41	42	46	43	44
°F Range	34-40	34-36	36-42	33-36	36-39	33-46	36-46	43-48	42-46	43-46
Conductivity, Av.	800	856	945	955	816	1090	1233	--	--	--
micromhos Range	740-940	740-980	760-1320	940-980	760-880	760-1420	800-1900	--	--	--
Effluent Rate, Av.	348	368	205	205	120	267	220	207	166	122
gpm Range	348-395	320-395	165-205	180-340	--	170-267	--	--	150-195	--
Turbidity, Av.	2	5	8	5	5	6	9	7	20	18
jtu Range	1-5	2-7	6-10	5-6	5-6	5-8	6-12	4-11	8-26	5-31

Table 24. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE PROCESS CONDITIONS (PROCESS NO. 11)

Water source	Average flow, gpm	Run duration hrs	Water treated gal	Test No.
Proctor No. 2	202	7.0	84,880	35-71
	253	22.3	338,610	37-71
	256	14.8	227,445	36-71

Water Flow:

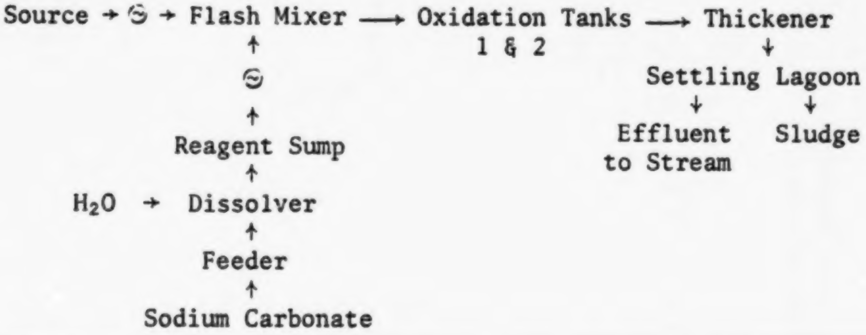
Source
Flash Mixer
Oxidation Tanks
1 & 2
Thickener
Settling Lagoon
Effluent to Stream
Sludge
Reagent Sump
H_2O
Dissolver
Feeder
Sodium Carbonate

Table 25. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE RAW WATER ANALYSES

Test No.	36-71	37-71	35-71
Water Source	Proctor No. 2	Proctor No. 2	Proctor No. 2
Temp., °F, Av.	50	50	50
pH, Av.	3.45	3.50	3.57
Range	3.20-3.92	3.43-3.69	3.83-3.95
Acidity, Av.	1167	1121	1386[a]
mg/l Range	1104-1259	---	---
Fe II, Av.	149	159	252[a]
mg/l Range	143-154	---	---
Fe III, mg/l, Av.	111[a]	101[a]	21[a]
Al, mg/l, Av.	233[a]	223[a]	240[a]
Ca, mg/l, Av.	201[a]	201[a]	201[a]
Mg, mg/l, Av.	97[a]	97[a]	97[a]
So_4, mg/l, Av.	1542	1748	1938
Conductivity, Av. micromhos	2700	2740	2700

[a] Values established from monthly analyses.

Table 26. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE - PROCESS AND REAGENT REQUIREMENTS

Test No.	36-71	37-71	35-71
Water Source	Proctor No. 2	Proctor No. 2	Proctor No. 2
Influent Fe II			
lbs.	281.6	448.0	178.3
lbs./hr.	19.090	20.132	25.476
Fe II Reacted			
lbs.	229.6	448.0	178.3
Minimum Oxidation Rate			
mg/l/hr.	121.0	158.5	251.8
Theoretical Reagent Required			
lbs.	2345.8	3357.1	1039.7
lbs./hr.	158.50	150.54	148.53
lbs./1,000 gal.	10.31	9.91	12.25
Reagent Utilized			
lbs.	5230	9787	2330
lbs./hr.	353.38	438.88	332.86
lbs./1,000 gal.	22.99	28.90	27.45
Alkali Consumption Utilized/Theoretical			
%	222.6	291.5	224.1
Retention Time (minutes)			
Flash Mixer	5.8	5.9	7.3
Oxidation Tanks	77.6	78.5	98.3
Theoretical Oxygen Requirement			
lbs.	40.3	64.2	25.5
lbs./hr.	2.735	2.884	3.649
lbs./1,000 gal.	0.1772	0.1900	0.3004
ft.^{3} (std.)	452.6	720.1	286.7
ft.^{3}/hr. (std.)	30.69	32.36	40.96
Oxygen Utilized as Supplied by Aerators			
Maximum			
lbs.	33.3	64.2	27.2
lbs./hr.	2.255	2.884	3.892
lbs./1,000 gal.	0.1464	0.1900	0.3204
Minimum			
lbs.	14.0	35.5	20.0
lbs./hr.	0.951	1.595	2.862
Oxygen Transfer Rate Based on Retention Time			
ppm O_2/water	22.8	29.8	63.1
ppm/hr. O_2/water	17.6	22.8	38.5
Oxygen Transferred as Required by Iron			
%	121.3	100.0	93.8
Oxygen Transferred as Aerator Rating			
%	3.8	4.8	6.5
Sludge Production			
gal./1,000 gal. treated[1]	(110)	(110)	(110)
solids content, mg/l	7000	7500	5550
lbs. dry solids/ 1,000 gal. treated	6.43	6.87	5.10

[1] Two hour settling time.

Table 27. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE OXIDATION TANK ANALYSES (PROCESS NO. 11)

Test No.	36-71	37-71	35-71
Water Source	Proctor No. 2	Proctor No. 2	Procotor No. 2
ph, Av.	6.83	7.39	7.53
Range	5.58-8.08	6.03-8.89	7.05-7.95
Acidity or Alkalinity[a]			
mg/l, Av.	-36	-914	-406
Range	-634-+293	---	---
Fe II, Av.	-33	---	---
mg/l Range	19-56	---	---
Fe III, mg/l, Av.	1	---	1
Suspended Solids, mg/l, Av.	770	824	610
Dissolved Oxygen, Av.	0.2	---	2.3
ppm Range	---	---	2.0-2.7
Oxidation-Reduction Potential, milliv.			
Av.	206	242	176
Range	170-260	220-280	168-194

[a] Negative value is Alkalinity.

Table 28. TREATMENT OF COAL MINE DRAINAGE WITH SODIUM CARBONATE THICKENER EFFLUENT ANALYSES (PROCESS NO. 11)

Test No.	36-71	37-71	35-71
Water Source	Proctor No. 2	Proctor No. 2	Proctor No. 2
pH, Av.	7.60	7.71	7.56
Range	7.56-7.63	7.50-7.81	7.48-7.60
Temp., Av.	46	47	44
°F Range	45-46	46-49	42-46

Table 29. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED DOLOMITE - PROCESS CONDITIONS

Water Source	Average Flow, gpm	Run Duration hrs	Water Treated gal	Test No.
	Process No. 11			
Proctor No. 2	83	13.5	67,900	59-71
	105	11.3	71,700	58-71
	106	21.3	135,740	60-71

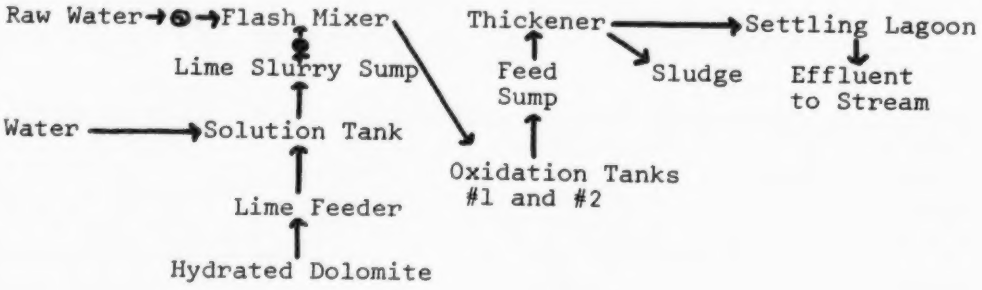
Raw Water
Flash Mixer
Lime Slurry Sump
Water
Solution Tank
Lime Feeder
Hydrated Dolomite
Oxidation Tanks #1 and #2
Feed Sump
Thickener
Sludge
Settling Lagoon
Effluent to Stream

Table 30. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED DOLOMITE RAW WATER ANALYSES

Test No.	59-71	58-71	60-71
Water Source	Proctor No. 2	Proctor No. 2	Proctor No. 2
Temp. °F, Av.	53	53	52
Range	53-54	52-54	52-53
pH, Av.	2.87	2.84	2.91
Range	2.83-2.90	2.83-2.84	2.90-2.91
Acidity, Av.	1352	1365[a]	1413
mg/l Range	1331-1364		
Fe II, mg/l, Av.	266	305[a]	305
Fe III, mg/l, Av.	>0	>0	>0
Al, mg/l, Av.	191[a]	191[a]	191[a]
Ca, mg/l, Av.	158	149[a]	149
Mg, mg/l, Av.	62	61[a]	61
SO_4, mg/l, Av.	1773	1937[a]	1793
Conductivity, Av.	3777	3780	3802
micromhos	3750-3790	3760-3790	3800-3810

[a] Values established from monthly analyses.

Table 31. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED DOLOMITE - PROCESS AND REAGENT REQUIREMENTS

Test No.	59-71	58-71	60-71
Water Source	Proctor No. 2	Proctor No. 2	Proctor No. 2
Influent Fe II			
lbs.	149.2	177.4	321.3
lbs./hr.	11.049	15.765	15.119
Fe II Reacted			
lbs.	148.9	177.4	321.3
Minimum Oxidation Rate			
mg/l/hr.	263.1	296.7	283.6
Theoretical Reagent Required*			
lbs.	540.6	576.2	1129.1
lbs./hr.	40.0	51.0	53.0
lbs./1,000 gal.	7.96	8.05	8.32
Reagent Utilized			
lbs.	504	840	1200
lbs./hr.	37.3	74.3	167.4
lbs./1,000 gal.	7.42	11.72	26.26
Alkali Consumption Utilized/Theoretical			
%	93.2	145.8	106.3
Alkali Consumption $\frac{\text{Theor. + Effl. Alk.}}{\text{Utilized}}$			
%	110	70	32
Retention Time (minutes)			
Flash Mixer	17.8	14.1	14.0
Oxidation Tanks	239	189	187
Theoretical Oxygen Requirement			
lbs.	21.4	25.4	46.0
lbs./hr.	1.583	2.258	2.166
lbs./1,000 gal.	0.3151	0.3542	0.3389
ft.^{3} (std.)	239.8	285.1	516.5
ft.^{3}/hr. (std.)	17.76	25.34	24.31
Oxygen Utilized as Supplied by Aerators			
Maximum			
lbs.	21.3	25.4	46.0
lbs./hr.	1.580	2.258	2.166
lbs./1,000 gal.	0.3137	0.3542	0.3389
Minimum			
lbs.	15.5	19.6	34.9
lbs./hr.	1.151	1.746	1.640
Oxygen Transfer Rate Based on Retention Time			
ppm O$_2$/water	151.4	135.1	127.2
ppm/hr. O$_2$/ water	38.0	42.9	40.8
Oxygen Transferred as Required by Iron			
%	100.2	100.0	100.0
Oxygen Transferred as Aerator Rating			
%	2.6	3.8	3.6
Sludge Production	(18		
gal./1,000 gal. treated	110 hours)	--	334
solids content, mg/l	13200	--	5800
lbs. dry solids/ 1,000 gal. treated	12.1	11.4	16.2

*Based on titre value of 81.03%.

Table 32. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED DOLOMITE OXIDATION TANK ANALYSES (PROCESS NO. 11)

Test No.	58-71	60-71
Water source	Proctor No. 2	Proctor No. 2
pH, Av.	7.51	8.03
Range	7.01-8.20	7.31-8.95
Acidity of Alkalinity[a], Av.	-28	-26
mg/l Range	---	-84-+36
Iron II, Av.	1	1
mg/l Range	---	---
Oxidation-Reduction Potential milliv., Av.	202	229
Range	192-214	220-242
Suspended Solids, Av.	1370	1940
ml/l Range	---	1870-1990

[a] Negative value is Alkalinity.

Table 33. TREATMENT OF COAL MINE DRAINAGE WITH HYDRATED DOLOMITE SETTLING LAGOON EFFLUENT ANALYSES

Test No.	59-71	58-71	60-71
Water source	Proctor No. 2	Proctor No. 2	Proctor No. 2
pH, Av.	7.48	7.52	7.54
Range	7.20-7.58	7.51-7.53	7.50-7.57
Alkalinity, Av.	-36	---	---
mg/l Range	-47--28	---	---
Iron II, Av. mg/l	1	---	---
Calcium, Av. mg/l	580	---	---
Magnesium, Av. mg/l	53	---	---

Table 34. TREATMENT OF COAL MINE DRAINAGE WITH PEBBLED DOLOMITE
RAW WATER ANALYSES (PROCESS NO. 11)

Test No.	61-71	62-71	63-71	64-71
Water source	Tyler Run	Bennett's Branch	Proctor No. 1	Proctor No. 2
Temp. °F, Av.	52	61	55	53
Range	48-62	60-65	52-60	53-56
pH, Av.	3.60	3.54	2.81	2.39
Range	3.51-4.00	3.20-3.81	2.80-2.82	2.38-2.39
Acidity, Av.	102	119	438	1395
mg/l Range	6-188	45-170	429-446	1336-1474
Fe II, Av.	4	18	69	310
mg/l Range	1-16	13-22		308-311
Fe III, Av. mg/l	>0	>0	>0	>0
Al, mg/l, Av.	12[a]	26[a]	86[a]	192[a]
Ca, mg/l, Av.	40	34	96	153
Range	31-58	31-37		
Mg, mg/l, Av.	17	11.5	76[a]	65
Range	12-25	11-12	---	---
SO_4, mg/l, Av.	169	198	726[a]	1869
Range	133-192	191-204	---	1855-1883
Conductivity, Av.	542	631	1540	4275
micromhos Range	500-600	590-690	600-1800	900-5400

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[a] Values established from monthly analyses.

Table 35. TREATMENT OF COAL MINE DRAINAGE WITH PEBBLED DOLOMITE - PROCESS AND REAGENT REQUIREMENTS

Test No.	61-71	62-71	63-71	64-71
Water Source	Tyler Run	Bennett's Branch	Proctor No. 1	Proctor No. 2
Influent Fe II				
lbs.	13.3	25.3	24.1	211.0
lbs./hr.	0.122	0.559	1.071	13.187
Fe II Reacted				
lbs.	10.0	23.9	23.4	208.3
Minimum Oxidation Rate				
mg/l/hr.	3.0	16.8	65.9	302.8
Theoretical Reagent Required[1]				
lbs.	206.5	101.6	93.5	577.2
lbs./hr.	1.89	2.24	4.16	36.08
lbs./1,000 gal.	0.51	0.60	2.20	7.00
Reagent Utilized				
lbs.	207	189	565	552
lbs./hr.	1.90	4.17	25.11	34.50
lbs./1,000 gal.	0.51	1.11	13.27	6.69
Alkali Consumption Utilized/Theoretical				
%	100.2	186.0	604.3	95.6
Retention Time (minutes)				
Flash Mixer	24.3	23.9	47.9	17.4
Oxidation Tanks	326	320	641	234
Theoretical Oxygen Requirement				
lbs.	1.9	3.6	3.5	30.2
lbs./hr.	0.018	0.080	0.153	1.889
lbs./1,000 gal.	0.0047	0.0211	0.0822	0.3661
ft.^{3} (std.)	21.4	40.7	38.7	339.2
ft.^{3}/hr. (std.)	0.20	0.90	1.72	21.199
Oxygen Utilized as Supplied by Aerators				
Maximum				
lbs.	1.4	3.4	3.4	29.9
lbs./hr.	0.013	0.076	0.149	1.865
lbs./1,000 gal.	0.0035	0.0200	0.0799	0.3624
Minimum				
lbs.	nil	nil	0.023	23.2
lbs./hr.	nil	nil	0.001	1.448
Oxygen Transfer Rate Based on Retention Time				
ppm O_2/water	2.3	12.8	102.6	170.8
ppm/hr. O_2/water	0.43	2.4	9.6	43.8
Oxygen Transferred as Required by Iron				
%	138.5	105.3	102.7	101.3
Oxygen Transferred as Aerator Rating				
%	0.02	0.1	0.2	3.1
Sludge Production				
gal./1,000 gal. treated	10	10	50	170
solids content, mg/l	8300	14200 (10 hrs)	9000	5380
lbs. dry solids/1,000 gal. treated	0.69	1.18	3.75	10.40

[1] Based on calcium content of 93.24% of theoretical.

Table 36. TREATMENT OF COAL MINE DRAINAGE WITH PEBBLED DOLOMITE OXIDATION TANK ANALYSES (PROCESS NO. 11)

Test No.	61-71	62-71	63-71	64-71
Water source	Tyler Run	Bennett's Branch	Proctor No. 1	Proctor No. 2
pH, Av.	7.44	7.63	7.89	7.87
Range	5.50-8.25	7.22-8.60	7.20-9.34	7.10-8.26
Acidity of Alkalinity[a] mg/1, Av.	-10	+3	+21	-35
Range	-61-+48	-17-+21	+4-+38	-149-+24
Iron II, Av.	1	1	2	4
mg/1 Range	---	---	1-2	1-27
Oxidation-Reduction Potential Milliv., Av.	293	242	197	238
Range	265-450	232-254	162-213	236-240
Suspended Solids, mg/1, Av.	83	142	---	1245
Range	4-150	102-196	---	1240-1250
Settling Lagoon Effluent Analysis				
pH, Av.	9.10	9.48	9.56	9.64
Range	7.53-9.78	8.73-9.80	9.28-9.71	9.60-9.70

[a] Negative value is Alkalinity.

Table 37. TREATMENT OF COAL MINE DRAINAGE WITH DOLOMITIC LIMESTONE - PROCESS CONDITIONS

Water Source	Average Flow, gpm	Run Duration hrs	Water Treated gal	Test No.
	Process No. 20			
Tyler Run	89	13.0	69,930	50-71
	Process No. 23a			
Tyler Run	136	24.0	196,880	51-71
	Process No. 23			
Bennett's Branch	45	17.5	47,650	52-71
	139	16.5	137,790	53-71
	Process No. 24			
Proctor No. 1	100	45.0	270,240	54-71
	78	30.5	144,120	55-71

Table 38. TREATMENT OF COAL MINE DRAINAGE WITH DOLOMITIC LIMESTONE - RAW WATER ANALYSES

Test No.	50-71	51-71	52-71	53-71	54-71	55-71
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1
Temp. °F, Av.	51	57	54	53	50	53
Range	49-55	47-64	53-56	50-56	47-56	49-58
pH, Av.	3.69	3.87	4.61	5.02	3.80	3.68
Range	3.45-3.80	3.81-3.90	3.87-5.00	4.85-5.13	3.50-5.13	3.61-3.71
Acidity, Av.	65	86	43	28	260	354
mg/1 Range	60-69	—	26-61	22-35	224-300	293-414
Fe II, Av.	1.5	1	6	5	43	65
mg/1 Range	1-2	--	5-8	--	19-54	56-72
Al, mg/1, Av.	12*	16*	16*	18*	80*	80*
Ca, mg/1, Av.	22*	23	16	16	48	68
Range	21-23	--	—	—	47-49	62-73
Mg, mg/1, Av.	9*	9	6	7	20	22
Range	--	--	--	6-7	19-20	19-25
SO_4,mg/1, Av.	138*	134	85	69	412	591
Range	135-140	133-135	--	59-81	399-424	588-624
Conductivity, Av.	475	383	270	252	1150	1496
micromhos Range	400-600	350-400	210-330	240-270	1060-1240	1400-1580

* Values established from monthly analyses.

Table 39. TREATMENT OF COAL MINE DRAINAGE WITH DOLOMITIC LIMESTONE - PROCESS AND REAGENT REQUIREMENTS

Test No.	50-71	51-71	52-71	53-71	54-71	55-71
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1
Influent Fe II						
lbs.	--	--	--	--	95.9	77.2
lbs./hr.	--	--	--	--	2.132	2.530
Fe II Reacted						
lbs.	--	--	--	--	93.7	98.4
Theoretical Reagent Required[1]						
lbs.	36.6	129.9	16.48	31.07	578.2	411.0
lbs./hr.	2.82	5.41	0.94	1.88	12.85	13.48
lbs./1,000 gal.	0.52	0.66	0.35	0.23	2.14	2.85
Reagent Utilized						
$CaCO_3 \cdot MgCO_3$						
lbs.	500	--	100	75.0	140	290
lbs./hr.	38.46	--	5.71	4.55	3.11	9.51
lbs./1,000 gal.	7.15	--	2.10	0.54	0.52	2.01
$CaCO_3$						
lbs.	--	111.0	--	--	501	1251
lbs./hr.	--	4.63	--	--	11.13	41.02
lbs./1,000 gal.	--	0.56	--	--	1.85	8.68
Total $CaCO_3 \cdot MgCO_3$ Equivalent[2]						
lbs.	--	102.0	--	--	601	1440
lbs./hr.	--	4.25	--	--	13.35	47.2
lbs./1,000 gal.	--	0.52	--	--	2.22	9.99
Alkali Consumption Utilized/Theoretical %	1366.1	78.5	606.8	241.4	103.9	350.4
Alkali Consumption $\frac{\text{Theor. + Effl. Alk.}}{\text{Utilized}}$ %	--	146	--	--	106	30
Retention Time (minutes)						
Flash Mixer	16.6	10.9	32.9	10.7	14.8	19.0
Oxidation Tanks	--	--	--	--	256	200

Table 39 (continued). TREATMENT OF COAL MINE DRAINAGE WITH DOLOMITIC LIMESTONE - PROCESS AND REAGENT REQUIREMENTS

Test No.	50-71	51-71	52-71	53-71	54-71	55-71
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1
Theoretical Oxygen Requirements						
lbs.	--	--	--	--	13.7	11.1
lbs./hr.	--	--	--	--	0.305	0.362
lbs./1,000 gal.	--	--	--	--	0.0507	0.0770
ft.^{3} (std.)	--	--	--	--	154.2	124.0
ft.^{3}/hr. (std.)	--	--	--	--	3.427	4.07
Oxygen Utilized as Supplied by Aerators						
Maximum						
lbs.	--	--	--	--	13.4	10.9
lbs./hr.	--	--	--	--	0.298	0.357
lbs./1,000 gal.	--	--	--	--	0.0496	0.0756
Sludge Production						
gal./1,000 gal. treated	[4]	[4]	[4]	--	(20)	(20)
solids content, mg/l	91000	34000	54000	--	70000	70000
lbs. dry solids/1,000 gal. treated	3.04	1.14	1.77	5.00	11.70	10.8

[1] Assumed to be 95.28% $CaCO_3 \cdot MgCO_3$.

[2] Based on limestone supplement only.

Table 40. TREATMENT OF COAL MINE DRAINAGE WITH DOLOMITIC LIMESTONE - LIMESTONE REACTOR EFFLUENT ANALYSES

Test No.	50-71	51-71			52-71	53-71	54-71[1]	55-71
Water Source	Tyler Run	Tyler Run			Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1
			After Limestone Addition	Settling Lagoon Influent				
pH, Av.	7.3	5.7	7.2	6.2	7.9	7.3	6.8	6.46
Range	7.2-8.5	5.4-6.0	6.5-7.7	6.1-6.4	7.1-8.0	7.2-7.48	5.02-7.0	4.99-7.30
Fe^{+2}, Av.	1	2	1	1	1	2	25	48
mg/1 Range	--	--	--	--	--	--	1-45	29-66
Alkalinity, Av.	39	13*	5	9*	10	15	15	91*
mg/1 Range	16-118	9*-17*	21-8*	--	0-17	17-17*	17-155*	51*-189*
Suspended Solids, mg/1	600	216	--	--	364	136	1300	--

*Acidity

[1] Could not maintain Fe levels even with $CaCO_3$ addition. Needed longer reaction time and oxidation. Put oxidation tanks in circuit.

Table 41. TREATMENT OF COAL MINE DRAINAGE WITH DOLOMITIC LIMESTONE AND PULVERIZED LIMESTONE OXIDATION TANK ANALYSIS (PROCESS NO. 24)

Test No.	54-71	55-71
Water source	Proctor No. 1	Proctor No. 1
pH, Av.	7.76	7.21
Range	7.50-8.05	6.50-7.70
Acidity or Alkalinity[a] mg/l, Av.	-22	-1
Range	-36--17	-17-+41
Iron II, Av. mg/l	1	4
Range	---	1-20
Iron III, mg/l, Av.	1	1
Suspended Solids, Av. ml/l	571	775
Range	550-592	761-788

[a] Negative value is Alkalinity.

Table 42. TREATMENT OF COAL MINE DRAINAGE WITH DOLOMITIC LIMESTONE - SETTLING LAGOON EFFLUENT ANALYSES

Test No.	50-71	51-71	52-71	53-71	54-71	55-71
Water Source	Tyler Run	Tyler Run	Bennett's Branch	Bennett's Branch	Proctor No. 1	Proctor No. 1
pH, Av.	7.29	7.25	7.47	7.47	7.50	7.54
Range	7.29-7.30	6.90-7.50	7.42-7.51	7.42-7.52	7.00-7.60	7.28-7.63
Alkalinity, Av.	--	-13	--	--	-28	-13
mg/1 Range	--	-18 - -8	--	--	-52 - +1	--
Iron II, Av.	--	--	--	--	2	--
mg/1 Range	--	--	--	--	0-5	--
Iron, Total, Av., mg/1	--	--	--	--	3	1
Suspended Solids, Av.	--	24	--	--	11	9
mg/1 Range	--	--	--	--	10-11	--
Calcium, Av., mg/1	--	54	41	55	76	--
Magnesium, Av., mg/1	--	13	7	11	14	--

APPENDIX D

PROCEDURES FOR THE ANALYSIS OF COAL MINE DRAINAGE WATERS

The methods employed to analyze coal mine drainage waters for its various components and properties vary widely depending upon the purposes and history of the particular laboratory. Since some constituents in these waters are arbitrarily defined, minor procedural variation can change the results significantly. The acidity determination is especially subject to these discrepancies. Most commonly, the base of the procedures employed has been "Standard Methods for the Examination of Water and Waste Water, American Public Health Association."[86] These analytical methods are widely used and commonly accepted procedures which have been adapted to this application. During the studies they were consistent with procedures employed by the Pennsylvania Department of Environmental Resources, formerly the Department of Health.[87] The U.S. Environmental Protective Agency--through its former agency, the Federal Water Quality Administration--issued a manual of analytical procedures[88] which has recently been updated[89] but it does not specify procedures for coal mine drainage for some components. The details of the procedures herein described are available in report form.[90]

SAMPLE COLLECTION

Samples were collected, transported and stored by recognized procedures to be representative of the materials being collected and to minimize change prior to analysis. Water samples containing hydrolyzable salts are especially subject to change.

Two types of special stream samples were collected. One was used to determine ferrous iron content and the other to determine total iron, aluminum, calcium, and magnesium content. Plastic bottles with a capacity of 100 milliliters are employed for the collection of 25 ml samples by pipette. For the sample to determine ferrous iron content, ten milliliters of a phosphoric-sulfuric acid mixture is added to the container prior to going into the field. This acid mixture is prepared as follows: mix 150 ml of sulfuric acid (1.84 sp. gr.) and 150 ml of phosphoric acid (1.75 sp. gr.) with enough deionized-distilled water to measure 1000 ml. This particular reagent is employed to prevent or minimize the oxidation of the ferrous iron until the sample can be analyzed.

Prior to collecting the sample to be used in determining the total iron, aluminum, calcium and magnesium content, two milliliters of concentrated HCl is added to the container to prevent the hydrolysis of either ferric iron or aluminum sulfate.

DETERMINATION OF pH

Determination of pH is carried out as soon as possible after a sample has been returned to the laboratory. Change which occurs in a sample as it stands prior to measurement could be related to modification in carbon dioxide content, iron II oxidation, or hydrolysis of metals salts. Analysis of the sample is made electrometrically using a pH meter with a glass electrode and a reference potential provided by a saturated calomel electrode. In continuous plant measurements, a coating of the electrodes with a gypsum-iron hydroxide film can be a serious problem. Cleaning of the electrodes with 1:2 hydrochloric acid has been adequate to keep them responsive (86d); the frequency of cleaning could vary with the systems and electrode locations.

DETERMINATION OF CONDUCTIVITY

Conductivity has significance in regard to the kind and number of ionic species present. It may also be employed to check other analyses. Such procedures have been detailed (86g) including a table of conductance factors for ions commonly found in waters. This parameter is especially useful for indicating changes in ionic concentration. In this work the value was measured with an Industrial Instruments, Inc., Conductivity Bridge, Model RC16B2.

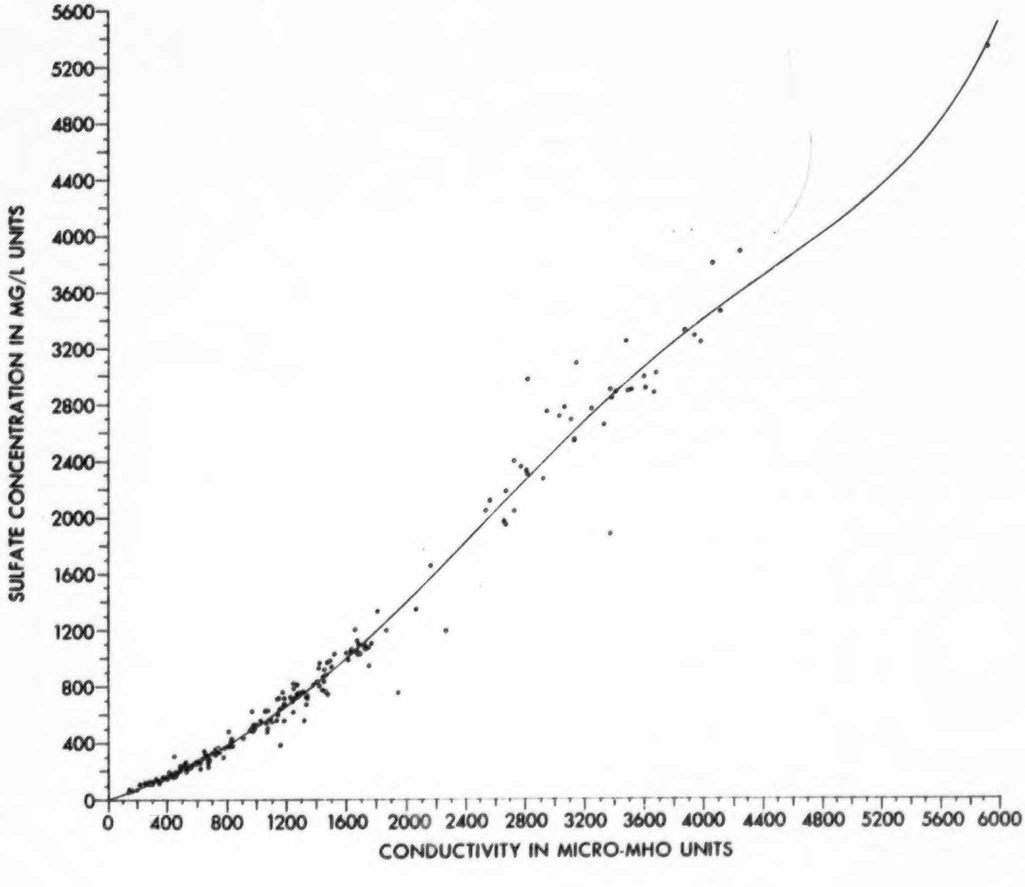
An interesting relationship, based upon monthly stream analyses, was observed between conductivity and sulfate concentration. These data are presented in Figure 1. On this plot, a line is given based upon Equation 10, developed by least squares fit as a fifth degree polynomial expression. This relationship can be helpful for control and other relative purposes due to the simplicity of conductivity determinations. The average error is 6.5 percent. This relationship applies for all of the Hollywood waters when conductivity is determined on a fresh sample.

The conductivity of mine drainage tends to increase upon standing due to iron II oxidation and oxide precipitation with subsequent increase in hydrogen ion concentration and its higher conductivity response. After standing about three months, a sample evaluated by this relationship yields sulfate values averaging about 18 percent high.

$$\begin{aligned} SO_4 \text{ (mg/l)} = {} & -4.4293552 + 0.43640856Y + 2.752911 \times 10^{-6}Y^2 \\ & + 1.8950735 \times 10^{-7}Y^3 - 6.3826551 \times 10^{-11}Y^4 \\ & + 5.9328758 \times 10^{-15}Y^5 \qquad (10) \end{aligned}$$

where Y = conductivity as micromhos at 20°C

Figure 1. Correlation between conductivity and sulfate concentration in mine drainage waters



DETERMINATION OF ACIDITY-ALKALINITY

Coal mine drainage is a mineralized or brackish water containing ions subject to complex asssociations and equilibria. These components make the acidity-alkalinity determination arbitrary. Existing procedures are not fully adequate, but should relate to the environmental conditions experienced or be otherwise defined.

The acidity (86a) of a water is defined as the capacity of that water to donate protons. This includes the non-ionized portions of weakly ionizing acids such as carbonic acid, acid salts, and hydrolyzable salts, as ferrous, ferric, and aluminum sulfate. When the sample has a low pH value, mineral acids contribute to the acidity. The term, acidity, as applied in these mine drainage studies expresses in stoichiometric equivalence the amount of alkali necessary to develop a stable condition within the water having a pH of 8.3 (phenolphthalein end point). Possible complex equilibria conditions, which will vary with pH, Eh and temperature, appear to function as a buffer, yielding an environment inconsistent with a "stable condition"[91]. According to Pyne and Yeates[92], acidity values may be high due to reactions with magnesium, aluminate, and manganous ions. These potential difficulties were previously cited by Lovell[64].

The alkalinity (86b) of a water is the capacity of that water to accept protons. Alkalinity is usually imparted by the bicarbonate, carbonate, and hydroxide components. In many natural waters, the alkalinity (and acidity) values are influenced by an equilibrium between these constituents.

In addition to the arbitrary basis of acidity-alkalinity evaluation, an additional uncertainty in the values from treated water samples has been related to sample treatment prior to analysis. There is evidence that samples may contain fine, insoluble, alkaline matter which will react during the analysis. Thus, an acidity-alkalinity value can vary depending upon the prior settling or filtration history of the sample.

The procedure for alkalinity-acidity employed is presumed to give results which are net value between acidity and alkalinity. For samples whose pH is below 4.5, the acidity is determined by adding hydrogen peroxide, heating to boiling to oxidize the iron, and followed with titration of the hot solution to a phenolphthalein end point with a standard base. When the sample has an initial pH above 4.5, the sample is first treated with a known volume of a standard acid solution and, following the steps just listed, back titrated with a standard base. The resulting value will express a net acidity or alkalinity.

DETERMINATION OF SOLUBLE SULFATES

Soluble sulfate content is determined gravimetrically by precipitation of barium sulfate (86f).

DETERMINATION OF SUSPENDED SOLIDS

Suspended solids values are established on a weight per unit volume basis developed by membrane filtration[86]. In samples containing large amounts of suspended solids, as thickener or settling lagoon underflow, quantitative filter paper is employed rather than filter membranes. The solids are normally dried at 105°C, although this temperature is increased to 180°C if concern exists regarding the presence of chemically-bound water in the crystals.

DETERMINATION OF DISSOLVED SOLIDS

To determine dissolved solids content, a sample is evaporated from a platinum vessel in a sand bath and finally dried at 105 or 180°C.

DETERMINATION OF TURBIDITY

Water turbidity is expressed in terms of arbitrary units rather than in terms of actual suspended material as weight per unit volume. It is evaluated by some light scattering procedure[86]. The actual suspended material is the more useful concept in mine drainage waters treatment and control although turbidity measurement is more rapid, convenient and may be adapted for continuous evaluation. A correlation between turbidity and suspended solids as well as the range of limitations of a given turbidity procedure must be established for each system considered.

A continuously-reading device functioning along these principles, and called "Surface Scatter Turbidimeter" (Hach Chemical Company) was employed satisfactorily in attaining plant data for this report.

DETERMINATION OF SPECIFIC GRAVITY

Mine Drainage Waters

The specific gravity of samples of mine drainage waters was determined at 20°C utilizing a hydrometer. The particular instrument employed, A.H. Thomas Urinometer (Squibb), catalog No. 3512, covered the range 0.0000 to 1.0600 and was readable to ±0.0001 units.

Sludge Samples

A standard 50 ml pycnometer, Gay-Lussac, ASTM, glass (Fisher Scientific Company, catalog No. 3-2400), was employed to determine the specific gravity of sludge samples. Total suspended solids are determined first. Dried sludge samples may be evaluated by this procedure utilizing boiled, deionized-distilled water as the displacement liquid. This liquid is probably less than satisfactory for this purpose due to solubility of some sludge components. Ethanol or acetone would likely be a more appropriate displacement liquid. A more convenient procedure applicable to "fluid" sludges simply established the weight of a given volume of sludge (50 or 100 ml), taking care to expel any trapped air bubbles.

DETERMINATION OF IRON

Ferrous and ferric iron content are determined by dichromate titration using diphenylamine sulfonate as an indicator. Total iron is determined by prior iron reduction with stannous chloride. When the iron content is less than five ppm, this procedure is insensitive and, for such samples, it is replaced by a colorimetric procedure (86c), utilizing the absorption of the iron - 1,10-phenanthroline complex at 508 millimicrons.

DETERMINATION OF ALUMINUM

Aluminum content is determined gravimetrically by difference involving separation of the R_2O_3 group and then subtracting the iron concentration expressed as the oxide.

DETERMINATION OF CALCIUM

Calcium content is determined gravimetrically by precipitation of calcium oxalate and its subsequent ignition to the oxide.

DETERMINATION OF MAGNESIUM

Magnesium content is usually determined on the filtrate from the calcium determination gravimetrically by precipitation of the magnesium ammonium phosphate (86a).

DETERMINATION OF METALS AND SULFATES IN SLUDGES

To determine metal and sulfate content in sludge, the dried sludge was solubilized with diluted hydrochloric acid and if necessary, any insolubles were fused. The cations and sulfates were determined by the procedures cited for water samples. Specialized procedures utilizing atomic absorption relied on lithium metaborate fusion procedures[93]. Spectrochemical procedures employed were detailed by O'Neil and Suhr[94].

A special series of analytical procedures were developed by Bodkin for these samples and this project. They involved a series of determinations from a single sample breakdown. The aluminum, iron, calcium, and magnesium were determined by chelometric titrations and back titrations with EDTA and CDTA. Magnesium and iron are obtained by "difference" in a series of titrations in which the calcium and aluminum are determined directly. Sodium is determined by atomic absorption. The details of these procedures will be published elsewhere[95].

DETERMINATION OF MANGANESE

The standard absorption procedure utilizing potassium periodate as an oxidant was employed [61] to determine manganese content.

APPENDIX E

ESTIMATING IRON OXIDATION RATES

Table 1. COMPUTER PROGRAM FOR ESTIMATING BIOCHEMICAL IRON OXIDATION RATES

```
FORTRAN IV G LEVEL  19              MAIN              DATE = 73342        13/10/05              PAGE 0001

0001            1 FORMAT (F6.0,1X,F5.1,1X,F5.1,1X,F5.1,1X,I3,1X,F4.1,1X,I3)              FE2    04
0002            2 FORMAT ('1', 'TO START',1X,'-',2X,'CAPACITY OF TANK IN GAL -- ',     FE2    05
                 CF6.0,6X,'1ST FE2 IN TANK -- ',F5.1,1X,'PPM',4X,'FLOW -- ',F5.1,      FE2    06
                 C1X,'GPM',4X,'FE2 INPUT -- ',F5.1,' PPM')                              FE2    07
0003            3 FORMAT (' ',11X,'LBS FE2 IN TANK -- ',F9.6,4X,'LBS FE2 INPUT -- ',FE2    08
                 CF8.6,1X,'PER INTERVAL')                                               FE2    09
0004            4 FORMAT (' ',F10.6,1X,'LBS',4X,F5.1,1X,'PPM',4X,F8.6,1X,'LBS CHANGEFE2    10
                 C',4X,F9.6,1X,'LBS GAIN',4X,I3,1X,'MIN')                               FE2    11
0005            5 FORMAT ('0',T50,'OX RATE -- ',F6.1,1X,'PPM/HR',2X,'OR',2X,F8.6,      FE2    12
                 C 1X,'LBS/INTERVAL')                                                   FE2   125
0006              DIMENSION ACAP(1),AFE1(1),AFLOW(1),AFEIN(1),IMIN(1),OXYRAT(1)         FE2    13
                 C,INTRVL(1)                                                            FE2   135
            C     READ IN RUN CONSTANTS                                                 FE2    14
            C                                                                           FE2    15
0007              READ (5,1,END=5000) ACAP(1),AFE1(1),AFLOW(1),AFEIN(1),IMIN(1)         FE2    16
                 C,OXYRAT(1),INTRVL(1)                                                  FE2   165
            C     FIGURE LBS FE2 IN TANK INITIALLY AND LBS FE2 PER MIN COMING IN        FE2    17
            C                                                                           FE2    18
0008              STRTLB=ACAP(1) * 8.3453 * AFE1(1) / 1000000.                          FE2    19
0009              AKEPLB=STRTLB                                                         FE2   195
0010              AINT=INTRVL(1)                                                        FE2   197
0011              FEINLB=AFLOW(1) * 8.3453 * AFEIN(1) / 1000000.*AINT                   FE2    20
0012              OXLOSS=-(OXYRAT(1)/1000000./60.*(ACAP(1)+AFLOW(1)*AINT)*8.3453        FE2   205
                 C*AINT)                                                                FE2   207
            C                                                                           FE2    21
            C     WRITE RUN CONSTANTS AND INITIAL CONDITIONS                            FE2    22
            C                                                                           FE2    23
0013              WRITE (6,2,ERR=6000) ACAP(1),AFE1(1),AFLOW(1),AFEIN(1)                FE2    24
0014              WRITE (6,3,ERR=7000) STRTLB,FEINLB                                    FE2    25
0015          750 WRITE (6,5,ERR=9000) OXYRAT(1),OXLOSS                                 FE2    26
0016              MIN=0                                                                 FE2    27
0017              ADELFE=0.                                                              FE2   273
0018              IMINIT=IMIN(1)                                                        FE2   275
0019              INTRV=INTRVL(1)                                                       FE2   277
0020              DO 500 I=INTRV,IMINIT,INTRV                                           FE2    28
0021              AINTLB=STRTLB+FEINLB+OXLOSS                                           FE2    29
0022              ANEWLB=AINTLB-(AFLOW(1)*AINT/(ACAP(1)+AFLOW(1)*AINT)*AINTLB)          FE2    30
0023              MIN=MIN+INTRV                                                         FE2    31
0024              ANWPPM=ANEWLB/ACAP(1)/8.3453*1000000.                                 FE2    32
0025              DELFE2=ANEWLB-STRTLB                                                  FE2    33
0026              DFEFLO=DELFE2-OXLOSS                                                  FE2   335
0027              WRITE (6,4,ERR=8000) ANEWLB,ANWPPM,DELFE2,DFEFLO,MIN                  FE2    34
0028              STRTLB=ANEWLB                                                         FE2    35
0029              ADELFE=ADELFE+ABS(DELFE2)                                             FE2   354
0030              IF (ADELFE-(.01*ANEWLB)) 500,500,300                                  FE2   355
0031          500 CONTINUE                                                              FE2    36
0032              STOP 1000                                                             FE2    37
0033          300 OXYRAT(1)=DFEFLO*1000000.*60./AINT/(ACAP(1)+AFLOW(1)*AINT)/8.3453 FE2   373
0034              STRTLB=AKEPLB                                                         FE2   374
0035              OXLOSS=-DFEFLO                                                        FE2   375
0036              GO TO 750                                                             FE2   376
0037         5000 STOP 5000                                                             FE2    38
0038         6000 STOP 6000                                                             FE2    39
0039         7000 STOP 7000                                                             FE2    40
0040         8000 STOP 8000                                                             FE2    41
0041         9000 STOP 9000                                                             FE2    42
0042              END
```

Table 2. EXAMPLE OF THE ESTIMATION OF BIOCHEMICAL IRON (II) OXIDATION RATES

```
TO START -  CAPACITY OF TANK IN GAL --  9931.    1ST FE2 IN TANK --  42.0 PPM    FLOW --  24.0 GPM    FE2 INPUT -- 351.0 PPM
            LBS FE2 IN TANK --  3.480839    LBS FE2 INPUT -- 0.351503 PER INTERVAL

  3.786587 LBS      45.7 PPM    0.305748 LBS CHANGE     0.305748 LBS GAIN     OX RATE --  0.0 PPM/HR  OR  0.0       LBS/INTERVAL
                                                                                 5 MIN

  3.484489 LBS      42.0 PPM    0.003651 LBS CHANGE                           OX RATE -- 43.7 PPM/HR  OR  -.305748 LBS/INTERVAL
  3.488096 LBS      42.1 PPM    0.003607 LBS CHANGE     0.309399 LBS GAIN        5 MIN
  3.491660 LBS      42.1 PPM    0.003564 LBS CHANGE     0.309355 LBS GAIN       10 MIN
  3.495181 LBS      42.2 PPM    0.003521 LBS CHANGE     0.309312 LBS GAIN       15 MIN
  3.498660 LBS      42.2 PPM    0.003479 LBS CHANGE     0.309269 LBS GAIN       20 MIN
  3.502098 LBS      42.3 PPM    0.003438 LBS CHANGE     0.309227 LBS GAIN       25 MIN
  3.505495 LBS      42.3 PPM    0.003397 LBS CHANGE     0.309186 LBS GAIN       30 MIN
  3.508851 LBS      42.3 PPM    0.003356 LBS CHANGE     0.309145 LBS GAIN       35 MIN
  3.512167 LBS      42.4 PPM    0.003316 LBS CHANGE     0.309104 LBS GAIN       40 MIN
  3.515444 LBS      42.4 PPM    0.003277 LBS CHANGE     0.309064 LBS GAIN       45 MIN
  3.518682 LBS      42.5 PPM    0.003238 LBS CHANGE     0.309025 LBS GAIN       50 MIN
                                                        0.308986 LBS GAIN       55 MIN

  3.481290 LBS      42.0 PPM    0.000451 LBS CHANGE                           OX RATE -- 44.2 PPM/HR  OR  -.308986 LBS/INTERVAL
  3.481736 LBS      42.0 PPM    0.000446 LBS CHANGE     0.309437 LBS GAIN        5 MIN
  3.482177 LBS      42.0 PPM    0.000441 LBS CHANGE     0.309432 LBS GAIN       10 MIN
  3.482613 LBS      42.0 PPM    0.000436 LBS CHANGE     0.309426 LBS GAIN       15 MIN
  3.483043 LBS      42.0 PPM    0.000430 LBS CHANGE     0.309422 LBS GAIN       20 MIN
  3.483468 LBS      42.0 PPM    0.000425 LBS CHANGE     0.309416 LBS GAIN       25 MIN
  3.483889 LBS      42.0 PPM    0.000421 LBS CHANGE     0.309411 LBS GAIN       30 MIN
  3.484303 LBS      42.0 PPM    0.000415 LBS CHANGE     0.309406 LBS GAIN       35 MIN
  3.484714 LBS      42.0 PPM    0.000410 LBS CHANGE     0.309401 LBS GAIN       40 MIN
  3.485119 LBS      42.1 PPM    0.000405 LBS CHANGE     0.309396 LBS GAIN       45 MIN
  3.485519 LBS      42.1 PPM    0.000401 LBS CHANGE     0.309391 LBS GAIN       50 MIN
  3.485915 LBS      42.1 PPM    0.000396 LBS CHANGE     0.309386 LBS GAIN       55 MIN
  3.486306 LBS      42.1 PPM    0.000391 LBS CHANGE     0.309381 LBS GAIN       60 MIN
  3.486692 LBS      42.1 PPM    0.000386 LBS CHANGE     0.309377 LBS GAIN       65 MIN
  3.487074 LBS      42.1 PPM    0.000381 LBS CHANGE     0.309372 LBS GAIN       70 MIN
  3.487451 LBS      42.1 PPM    0.000377 LBS CHANGE     0.309367 LBS GAIN       75 MIN
  3.487823 LBS      42.1 PPM    0.000373 LBS CHANGE     0.309362 LBS GAIN       80 MIN
  3.488192 LBS      42.1 PPM    0.000368 LBS CHANGE     0.309359 LBS GAIN       85 MIN
  3.488555 LBS      42.1 PPM    0.000363 LBS CHANGE     0.309354 LBS GAIN       90 MIN
  3.488914 LBS      42.1 PPM    0.000360 LBS CHANGE     0.309349 LBS GAIN       95 MIN
  3.489269 LBS      42.1 PPM    0.000355 LBS CHANGE     0.309345 LBS GAIN      100 MIN
  3.489620 LBS      42.1 PPM    0.000351 LBS CHANGE     0.309340 LBS GAIN      105 MIN
  3.489967 LBS      42.1 PPM    0.000347 LBS CHANGE     0.309337 LBS GAIN      110 MIN
  3.490310 LBS      42.1 PPM    0.000342 LBS CHANGE     0.309333 LBS GAIN      115 MIN
  3.490648 LBS      42.1 PPM    0.000339 LBS CHANGE     0.309328 LBS GAIN      120 MIN
  3.490983 LBS      42.1 PPM    0.000335 LBS CHANGE     0.309324 LBS GAIN      125 MIN
  3.491313 LBS      42.1 PPM    0.000330 LBS CHANGE     0.309320 LBS GAIN      130 MIN
  3.491639 LBS      42.1 PPM    0.000326 LBS CHANGE     0.309316 LBS GAIN      135 MIN
  3.491961 LBS      42.1 PPM    0.000322 LBS CHANGE     0.309312 LBS GAIN      140 MIN
  3.492280 LBS      42.1 PPM    0.000319 LBS CHANGE     0.309308 LBS GAIN      145 MIN
  3.492595 LBS      42.1 PPM    0.000315 LBS CHANGE     0.309304 LBS GAIN      150 MIN
  3.492906 LBS      42.1 PPM    0.000311 LBS CHANGE     0.309300 LBS GAIN      155 MIN
  3.493213 LBS      42.1 PPM    0.000307 LBS CHANGE     0.309297 LBS GAIN      160 MIN
  3.493517 LBS      42.2 PPM    0.000304 LBS CHANGE     0.309293 LBS GAIN      165 MIN
  3.493817 LBS      42.2 PPM    0.000300 LBS CHANGE     0.309290 LBS GAIN      170 MIN
  3.494114 LBS      42.2 PPM    0.000297 LBS CHANGE     0.309286 LBS GAIN      175 MIN
  3.494407 LBS      42.2 PPM    0.000293 LBS CHANGE     0.309282 LBS GAIN      180 MIN
  3.494697 LBS      42.2 PPM    0.000290 LBS CHANGE     0.309278 LBS GAIN      185 MIN
  3.494983 LBS      42.2 PPM    0.000286 LBS CHANGE     0.309276 LBS GAIN      190 MIN
  3.495265 LBS      42.2 PPM    0.000282 LBS CHANGE     0.309272 LBS GAIN      195 MIN
  3.495544 LBS      42.2 PPM    0.000279 LBS CHANGE     0.309268 LBS GAIN      200 MIN
  3.495820 LBS      42.2 PPM    0.000276 LBS CHANGE     0.309265 LBS GAIN      205 MIN
  3.496093 LBS      42.2 PPM    0.000273 LBS CHANGE     0.309261 LBS GAIN      210 MIN
  3.496362 LBS      42.2 PPM    0.000269 LBS CHANGE     0.309258 LBS GAIN      215 MIN
  3.496628 LBS      42.2 PPM    0.000266 LBS CHANGE     0.309255 LBS GAIN      220 MIN
  3.496891 LBS      42.2 PPM    0.000263 LBS CHANGE     0.309252 LBS GAIN      225 MIN
  3.497150 LBS      42.2 PPM    0.000259 LBS CHANGE     0.309249 LBS GAIN      230 MIN
  3.497407 LBS      42.2 PPM    0.000257 LBS CHANGE     0.309245 LBS GAIN      235 MIN
                                                        0.309242 LBS GAIN      240 MIN
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Table 3. EXAMPLE OF THE ESTIMATION OF AIR-IRON II OXIDATION RATES[a]

Water source	Fe_OII mg/1	Fe_FII mg/1	Al mg/1	pH	T°C	Rate (mg/1/hr)
Tyler Run	2	1	10	6	16	79
Bennett's Branch	15	7	30	6	16	172
Proctor No. 1	60	7	85	6	10	66
Proctor No. 2	400	7	200	6	10	180
Examples	5	0	1	5	10	150
	5	0	1	5	20	150
	5	0	1	7	20	156
	5	0	50	7	20	48
	50	7	1	5	10	10,500
	50	7	1	5	20	10,500
	50	7	1	7	20	10,920
	50	7	50	7	20	360
	125	7	1	5	10	16,140
	125	7	1	5	20	16,200
	125	7	1	7	20	16,800
	125	7	125	7	20	420
	450	7	1	5	10	31,560
	450	7	1	5	20	31,620
	450	7	1	7	20	32,820
	450	7	200	7	20	900

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[a] See Equation 6.

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